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Automated Control: A Review and Applications in Industrial Environmental Protection

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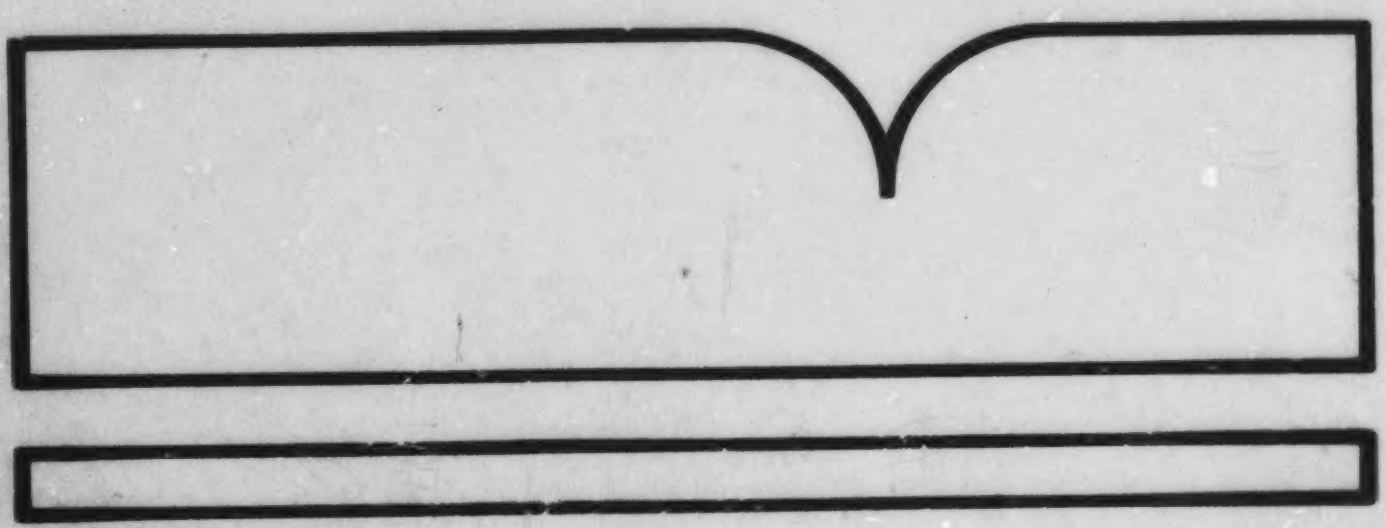

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AUTOMATED CONTROL: A REVIEW AND APPLICATIONS IN INDUSTRIAL ENVIRONMENTAL PROTECTION

by

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16. ABSTRACT

The report gives results of an examination of automatic control theory and its practical applications to environmental processes. A summary is given, emphasizing aspects of the theory that are likely to find application in optimizing environmental control systems. Several case studies, chosen based on the potential applicability of automatic control processes, are used to illustrate applications of several automatic control concepts. The basic equations are introduced in the development of a closed-loop transfer function for a blast furnace scrubber water recycle system. The mathematical complexities of handling a distributed parameter system are illustrated in a study of an acid gas removal system. The possibility of utilizing feedforward control is illustrated in an examination of fluidized-bed combustion with limestone control. Evaluations of various control options are considered within the context of a limestone scrubber slurry treatment system.

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Slurries	Stationary Sources	14B
Automatic Control	Acid Gas Removal	07A, 13I
Scrubbers		11F
Blast Furnaces		
Fluidized Bed Processes		
Calcium Carbonates		07B

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ABSTRACT

The objective of this study was to examine and summarize automatic control theory and its practical applications to pollution control processes. A summary is given emphasizing those aspects of the theory that are likely to find application in optimizing environmental control systems. The summary provided is intended to supply engineers evaluating pollution control technologies with an additional tool for planning and design. Several case studies, which have been chosen based upon the potential applicability of automatic control processes, have been used to illustrate applications of several automatic control concepts. The basic equations are introduced in the development of a closed-loop transfer function for a blast furnace scrubber water recycle system. The mathematical complexities of handling a distributed parameter system are illustrated in a study of an acid gas removal system. The possibility of utilizing feedforward control is illustrated in an examination of fluidized-bed combustion with limestone control, and evaluation of various control options are considered within the context of a limestone scrubber slurry treatment system.

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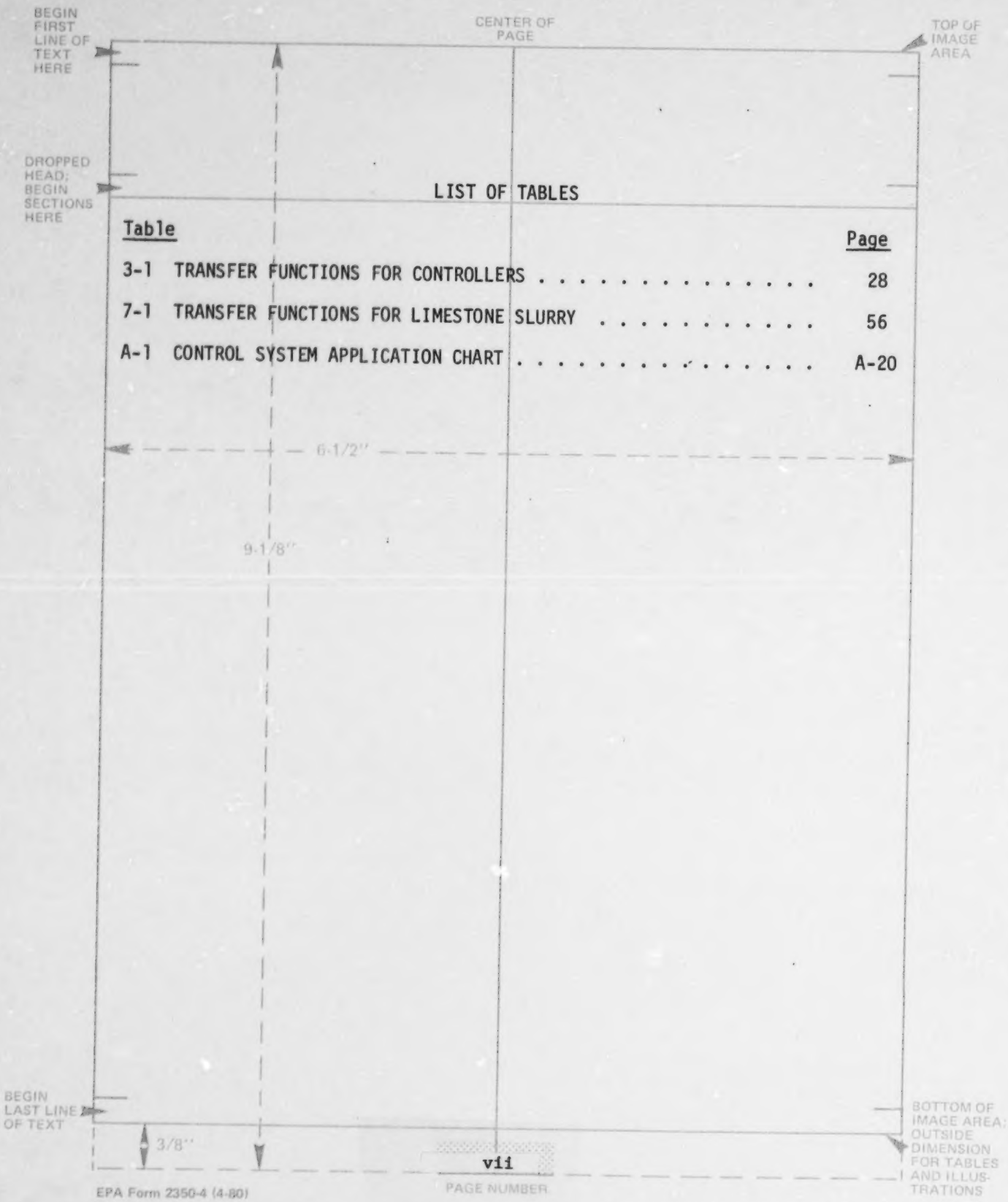
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1 INTRODUCTION

A control system has been defined as a system in which deliberate guidance or manipulation is used to achieve a prescribed value of a variable (ISA, 1979). A control system is generally designed to meet some performance specification at a minimum cost. Classical control system design techniques can be very effectively applied to the control of processes that are designed to reduce or eliminate potential environmental hazards associated with discharges from utilities or industrial plants. It should be noted that a control system is subdivided into a controlling system and a controlled system.

This report provides a review of the basic principles of control theory and demonstrates the applicability of the dynamic analysis techniques to energy and industrial pollution abatement processes. The information has been selected to guide an interested reader through some of the more important aspects of control theory and to enhance his appreciation for the analysis methods as they apply to many processes. Sections 2 and 3 provide an overview of control theory, including the sequential steps that comprise the approach to process control, technical terms, and control instrumentation and controllers.

Sections 4-7 present four case studies that demonstrate important technical aspects of control theory. The systems that were selected for examination in the case studies are all associated with technologies that were developed for environmental protection. The case studies are not intended to detail all aspects of automatic control for each of the subject systems. Collectively, however, the studies serve to illustrate the approach to control system analysis while pointing out certain limitations and complications in the practical application of control theory. Although simplified for the purpose of illustration, these studies progress through

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the sequence of steps necessary for evaluating an uncontrolled system and initiating the design of an effective system of control.

The appendix provides supplementary information regarding analysis approaches and reviews several topics related to system design and compensation.

Most industrial plants were run essentially manually prior to the 1940's, and many still rely almost exclusively on human control. In the late 1940's and 1950's, some feedback control was added to plants, but with little real consideration or appreciation for the actual process dynamics. Relative guides and experience were the only bases for automatic control system design. For the past two decades, however, dynamic analysis techniques and control theory (developed primarily by aerospace and electrical engineers) have been applied in the design of many industrial plant engineering systems (Luyben, 1973).

The practical engineer, with industrial experience, may question the need for dynamic analysis and sophisticated control system design techniques. It may be argued that probably 70 to 80 percent of the control loops in today's industrial plants are designed, installed, tuned, and operated quite successfully, by simple, rule-of-thumb, experience-generated techniques. Significantly, the remainder of the control system is likely to be designed through engineering based on control theory, and these control loops are often those most critical to successful plant operation.

Automatic process control has been applied only minimally to environmental control systems. In many cases, insufficient data on the performance of these systems are available for control system optimization. The fundamentals of operation must be verified before the advantages of automatic control can be effectively applied. It should be emphasized, however, that the same control techniques that have been demonstrated to optimize production could be equally effective for pollution control strategies.

Over the last two decades, an extensive mathematical methodology has been developed for dealing with time-varying behavior of physical systems when they are subjected to disturbances. When applied to chemical processes, the mathematical approach is called process dynamics (Buckley, 1964). The traditional areas of chemical engineering--thermodynamics,

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kinetics, unit operations, and transport phenomena--provide the foundation for the study of process dynamics. The mathematical models of process control are but extensions of the traditional unit operations equations to include time-dependent terms. Process dynamics is a vital element in control theory. Courses in process dynamics, control, and instrumentation are taught in many universities within the mechanical, chemical, and electrical engineering curricula.

The goals of a study of process dynamics and control should include:

1. Gaining an appreciation for the dynamics (i.e., the time-dependent responses) of engineering systems.

2. Learning to develop realistic but reasonable mathematical models of these systems, recognizing the advantages and pitfalls of simplifying assumptions.

3. Learning techniques for solving the differential equations that make up the mathematical model describing the transient response of important system variables.

4. Studying control system design/techniques--control strategy, stability theory, performance analysis, controller design, and simulation.

5. Becoming familiar with practical, commercial control equipment (the hardware that goes into a real control system--sensors, transmitters, controllers, and control valves).

A major source of information on instrumentation and control is the extensive headquarters library of the Instrument Society of America (ISA), which is located in the Research Triangle Park, North Carolina. ISA, which was established in 1945, is an international technical, scientific, and educational organization for engineers, scientists, managers, technicians, and students involved with instrumentation systems and automation. ISA has been responsible for establishing standardized terminology for process instrumentation. This standard terminology has been published by the American National Standards Institute (ANSI). The ISA publishes a monthly journal, (INTECH) Instrumentation Technology, and sponsors short courses on many aspects of instrumentation and control. Also, it has published several important textbooks.

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2. AUTOMATIC CONTROL SYSTEMS

Control theory deals with the dynamic response of a system to commands or disturbances (Hale, 1973). There are two parts in the application of control theory to be considered--dynamic analysis and control system design. The analysis phase establishes the output-input relationships that characterize a system. The design phase involves determining an appropriate control strategy and specifying control instrumentation to accomplish an acceptable response to perturbations on the system. Process control is merely a descriptive term applied to the control of chemical or similar physical processes.

2.1 APPROACH TO PROCESS CONTROL

For those control system problems that engineering experience and intuition cannot solve, the approach to process analysis and control system design is a consistent one involving the following sequence of steps:

1. Define a system and its components.
2. Establish performance criteria for the controlled system.
3. Formulate the mathematical transient model (i.e., examine process dynamics) of the uncontrolled system, and list the necessary assumptions.
4. Write the differential equations that describe the model.
5. Solve the equations for the desired output variable.
6. Examine the solution and the assumptions, and compare them with performance criteria.
7. Propose appropriate control strategy, and design the controlling system.
8. Perform a dynamic analysis of the complete system to assure stability.
9. Test the design by nonlinear simulation to confirm response.
10. Construct and test.

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To begin, one must obtain a quantitative mathematical model of the system. Since the systems under consideration are typically dynamic in nature, the descriptive equations are usually differential equations. Partial differential equations are involved in the modeling of distributed parameter systems (i.e., systems in which the continuous spatial distribution of a physical characteristic is taken into account). If the equations can be replaced by their linear series expansions, then the Laplace transform may be utilized in order to simplify the method of solution.

Other methods of solution, discussed in the appendix, may also be employed if necessary. Whether the equations describing the system are written as basically derived in the time domain, or as they have been transformed (e.g., Laplace transform), a characteristic equation is obtained from the differential equation(s) of a system. Analogous equations with similar solutions describe electrical, thermal, mechanical, and fluid systems. Because of the similarities in the defining equations, the solution of one system may be extended to all analogous systems with the same describing differential equations. In reality, the complexity of systems and the lack of definitions of all relevant dynamic factors frequently necessitate the introduction of simplifying assumptions concerning the system operation.

An appropriate control strategy and design will be based on the output-input response as determined by the dynamic analysis of the uncontrolled system and certain performance criteria. Factors to be considered are the required levels of accuracy, speed, cost, reliability, and ease of operation and maintenance. The operating environment must also be taken into account. Functional requirements for process control systems are: (1) stability; (2) reduction or elimination of effects of long-term disturbances; and (3) reduction or elimination of effects of short-term disturbances.

Nonlinear simulation of the controlled system is typically accomplished with an analog computer. Continuous variables are represented by voltages, and parametric values are controlled by potentiometers. Integrations are performed by high-gain amplifiers. A digital computer can also be programmed to perform simulation either alone or in conjunction with the analog computer.

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2.2 GENERAL CONCEPTS AND TERMINOLOGY OF CONTROL THEORY

Discussed below are some of the basic concepts and terms essential to an understanding of control theory.

2.2.1 Regulator and Servo Problems

Two basic problem types are encountered in process control--the regulator problem and the servo problem. In the regulator problem, the value of a particular variable is specified. The purpose of the control system is to maintain the level of that variable (called the controlled variable) at the specified value, in spite of changes in the system load. Load disturbances might be changes in throughput, feed conditions, or supply steam pressure. Most control systems for pollution abatement processes are of this type with the controlled variable being the level of a particular pollutant in a discharge stream.

A second type of problem, called the servomechanism-type (or servo) problem, involves tracking a variable according to prescribed set point changes. The set point is defined as "an input variable that sets the desired value of the controlled variable" (ISA, 1979). Set point changes can be made during startup of continuous processes or in batch processes. The function of the feedback controller in servo control is to drive the controlled variable to match the set point. This type of problem is encountered less frequently than the regulator problem in environmental applications of control theory.

2.2.2 Transfer Function

The transfer function is one of the most useful concepts in control theory. The transfer function of a linear equation describing system behavior is the ratio of the transform of the output variable (response function) to the transform of the input variable (forcing function). The transfer function, designated G(s), is defined by Equation (2-1).

$$G(s) = \frac{\text{Transform of output variable}}{\text{Transform of input variable}} = \frac{\text{Output}}{\text{Input}} \qquad (2\text{-}1)$$

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This definition holds for systems that are linear and stationary, with continuous variables and with zero initial conditions. Under these conditions, the transfer function can be expressed as a ratio of two polynomials in the complex Laplace variables. (See Section 2.2.4.)

The significance of the transfer function as defined by Equation (2-1) is that, if the transfer function of a system is known, then the characteristics of the behavior of the system itself are known. The denominator of Equation (2-1), Input(s), is called the characteristic function since it contains all of the physical characteristics of the subject system. The characteristic equation is obtained by setting the polynomial, Input(s), equal to zero. The roots of this equation determine the stability of the system and the general nature of the transient response. The numerator polynomial of Equation (2-1), Output(s), establishes the transient response as well as the steady-state value of the output. It should be kept in mind that the transfer function is an input/output description of the behavior of the system and does not include any information concerning the internal structure of the system and its behavior.

Either mathematical or experimental methods may be used to obtain the transfer function. In the experimental approach, a known input, typically a step function, is applied to the system, and the output is measured. The transfer function is then constructed from these data. Similarly, the transfer function may be obtained from operating data and curves.

As described, the mathematical approach is to transform the system equations (usually differential equations) and obtain the ratio of response to input. The transfer function for a complete system is obtained by the proper combination of the known transfer functions of the individual elements of the system. The transfer function is most useful for lumped parameter systems where transport lag may be neglected.

2.2.3 Block Diagrams

A signal flow block diagram is a useful, systematic way of writing the simultaneous differential and algebraic equations that describe the dynamic behavior of the components of a system. The diagram clarifies the relationship among the variables of these simultaneous equations.

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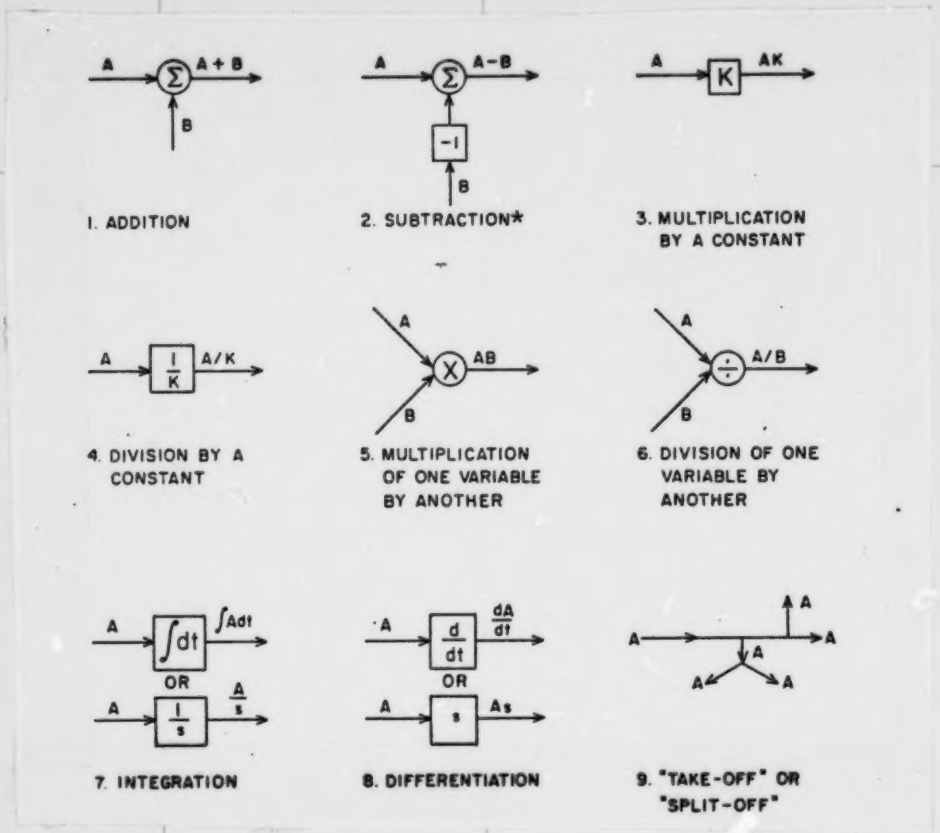
The basic element of the signal flow diagram is a block expressing a relation between two signals, the input and the output. The diagram may be used to represent the time domain relationship of the subject variables, or it may show the Laplace transform (or other transform) equivalent, i.e., the transfer function. The symbols that are commonly used to indicate the various mathematical operations in signal flow diagrams are shown in Figure 2-1 below.



*Often subtraction is signified in the same way as addition with a negative sign near the arrow tip at the intersection.

Figure 2-1. Symbols for signal flow block diagrams (reproduced from Buckley, 1964, with permission from the publisher, John Wiley and Sons, Inc.)

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2.2.4 The Laplace Transformation

The most common transformation encountered in control work is the Laplace transformation. It is primarily a method of tranforming linear ordinary differential equations with constant coefficients into algebraic equations. The chief advantages of using the Laplace transform for control studies are:

1. It permits rapid solutions for problems involving simultaneous linear differential equations with constant coefficients.
2. Initial boundary conditions are automatically incorporated in the solution for the response to any arbitrary forcing function.
3. In many cases, the nature of the response can be seen simply by inspecting the transformed equation. (Buckley, 1964).

In control work, the usual procedure is to Laplace transform the original equations that describe the time-varying or transient behavior of a physical system. Usually, the initial conditions are ignored. The transformed equations are then combined and rearranged to obtain the transfer function. Finally, the inverse transform is performed to obtain the desired solution.

The Laplace transform of a function f(t) is defined to be f(s) according to Equation (2-2).

$$f(s) = \int_0^\infty f(t)e^{-st}\,dt = L\,[f(t)]. \tag{2-2}$$

The principal motivation for applying the Laplace transform arises in connection with the time derivative which occurs frequently in the equations describing process dynamics. The operator L can be thought of as substituting multiplication by s for the operation of differentiation indicated in Equations (2-3) and (2-4).

$$L\left[\frac{df(t)}{dt}\right] = sf(s) - f(t)_{t=0} \tag{2-3}$$

$$L\left[\frac{d^n f}{dt^n}\right] = s^n f(s) - s^{n-1}f(0) - s^{n-2}f^{(1)}(0) - \ldots - sf^{(n-2)}(0) - f^{(n-1)}(0), \tag{2-4}$$

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where $f^{(i)}(0)$ indicates the i^{th} derivative of f(t) with respect to t, evaluated for t=0.

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2.3 CONTROL SYSTEM CONFIGURATIONS

In order to automate the control of a process, it is necessary to "feed" information to a controller which will, in turn, effect some change in a process variable, called the manipulative variable. The information received by the controller is data or a signal generated by a measuring device that senses fluctuations in some system parameter. In process instrumentation, the error is the algebraic difference between the indication of the measured signal and the ideal value (ISA, 1979).

The two basic configurations for control systems are feedback control and feedforward control. In feedback control, information from the output variable is "fed back" to an input manipulative variable. In feedforward control, the disturbance is detected in the input variable before it enters a process so that adjustments are made before the disturbance can propagate through the process. The basic control configurations are illustrated in Figure 2-2.

2.3.1 Feedback Control

Auto-control is usually associated with closed-loop, feedback control systems. The concept of using information about the deviation of a system from its desired state to control the system is called feedback control. In feedback control a measured variable is compared to its desired value to produce an actuating error signal which is acted upon in such a way as to reduce the magnitude of the error (ISA, 1979). The feedback concept is generally accepted to be the foundation for control system analysis and design.

The advantages of feedback control include:

1. A decrease in the sensitivity of the system to variations in the parameters of the process,
2. Ease of control and adjustment of the transient response of the system,
3. Improvement in the rejection of the disturbance and noise signals within the system,

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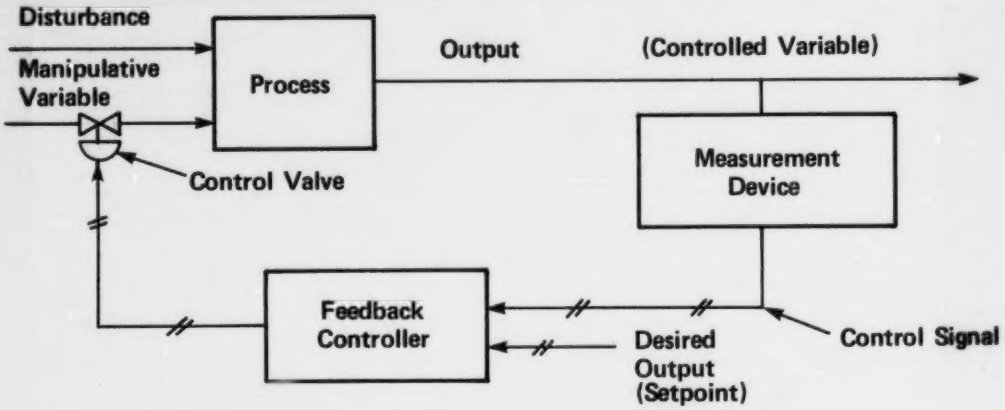
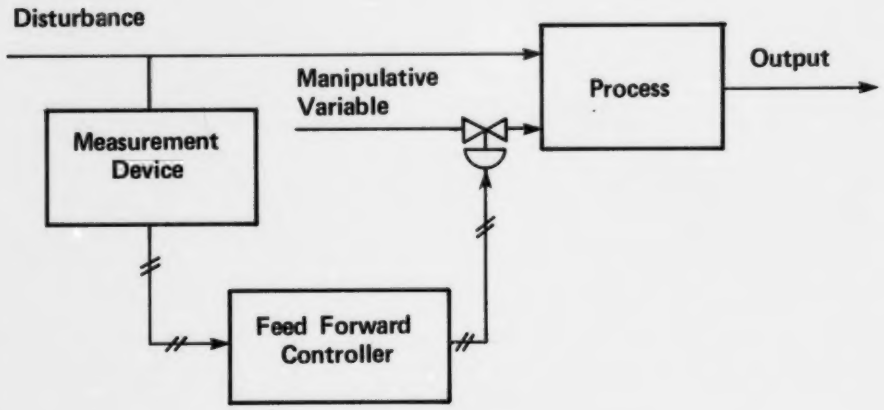


Feedback Control



Feedforward Control

Figure 2-2. Feedback and feedforward control

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4. Improvement in the reduction of the steady-state error of the system.

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There are, as might be expected, tradeoffs in the use of feedback control. Feedback reduces both the overall gain of a system (a disadvantage) and the errors associated with parameter changes and disturbance inputs and upsets (an advantage). The gain of a closed-loop system is expressed as the ratio of the output change to the input change at a specified frequency (ISA, 1979).

These two characteristics, which are common to all feedback control systems, can be demonstrated by consideration of the simple voltage amplifier of Figure 2-3a, where E_{in} is the control voltage (the input), and E_{out} is the amplified voltage (the output). Although electrical processes (amplifiers) and units (volts) are used for simplicity, the same considerations apply to chemical processes precipitators, scrubbers, etc.) and units (flow rates, temperatures, etc.). Neglecting the amplifier dynamics (a valid assumption for any well-designed or properly specified amplifier that might be used in a control system), then the open-loop input-output relationship is

$$E_{out} = AE_{in} \qquad \text{or} \qquad \frac{E_{out}}{E_{in}} = A$$

where A is the open-loop gain. If E_{in} is 1 volt, and E_{out} is equal to 100 volts, the system gain, A, is 100 volts/volt. If, however, there is a 10 percent error in A, so that A is 90 rather than 100, then E_{out} for the same 1 volt input will be only 90 volts or a 10 percent error in the output. In other words, there is a one-to-one correspondence between the amplifier characteristics and the output error.

Next, consider what happens in the block diagram (Figure 2-3b) when a fraction, β, of the output voltage is fed back and subtracted from the input voltage. The input-output relationship for this negative feedback system is found from simple algebra to be

$$\frac{E_{out}}{E_{in}} = \left(\frac{A}{1 + \beta A}\right) = \left(\frac{1}{1 + \beta A}\right) A \ .$$

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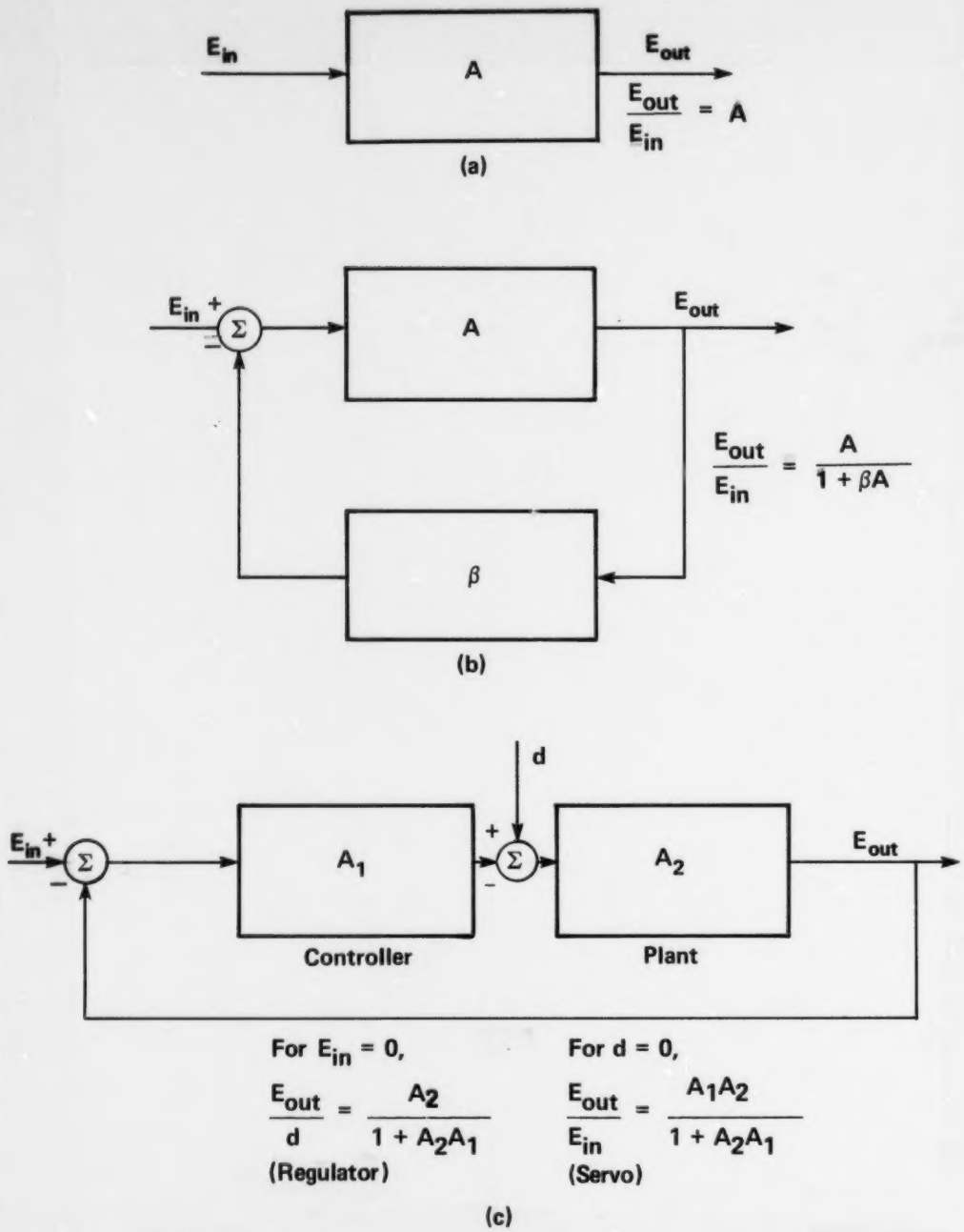

Figure 2-3. Input-output relationships for simple systems: (a) open-loop; (b) closed-loop; (c) closed-loop with disturbance input.

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The system gain has been reduced by the factor $1/(1+\beta A)$. If β is 0.1 and A is still 100 v/v, then the overall gain is 9.1 v/v, an order of magnitude reduction from the open-loop value of 100 v/v. This is an obvious disadvantage if an overall gain of 100 is to be maintained; additional amplifiers must be added to the forward path. However, now the same 10 percent error in the value of A results in an output error of less than 1 percent, which represents an order of magnitude reduction in the sensitivity of the system to changes in the amplifier gain.

The addition of the feedback loop has reduced both the gain of the system and its sensitivity to parameter changes in the forward path. It also has increased the complexity of the system. However, the sensitivity reduction not only tends to linearize any inherent nonlinearities in components but also allows the use of less accurate and less expensive hardware to achieve the same system accuracy as the corresponding open-loop system. The sensitivity reduction in the forward path has been accompanied by a shift to the feedback path of the one-to-one correspondence between system errors and changes in the feedback parameters. For example, a 10 percent error in β results in a 10 percent error in the output. Consequently, care must be taken in specifying the feedback path.

Analogous to the ways in which the feedback loop influences amplifier output, feedback control can also be used to reduce the errors arising from disturbance inputs entering a chemical process. In Figure 2-3c, the plant is represented by the gain A_2, the controller by the gain A_1, and the disturbance by d. Without the feedback loop and with A_2 equal to 10, the output error is simply 10 for a unity disturbance. With the feedback loop closed and only a disturbance input being considered (i.e., neglecting changes in the command input), with A_1 equal to 100, the error is dramatically reduced by two orders of magnitude to 0.01. The transfer functions for both the regulator response (disturbance input only) and the servo response (change in set point) are indicated on the figure.

2.3.2 Feedforward Control

The objective in feedforward control is to detect disturbances as they enter a process and make adjustments in manipulative variables so that

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output variables are held constant. The ISA (1979) defines feedforward control as "control in which information concerning one or more conditions that can disturb the controlled variable is converted, outside of any feedback loop, into corrective action to minimize deviations of the controlled variable." Feedforward control has received considerable emphasis in recent years. It offers the advantage of taking immediate action to accommodate a disturbance rather than detecting the disturbing influence after it has affected the controlled variable. In theory, feedforward control can result in the achievement of perfect control.

In practice, there are serious limitations to the use of feedforward control. The requirement of detecting the disturbance ahead of the process cannot always be met. Feed composition changes, for example, would have to be monitored by a continuous analyzer if feedforward control is to be used. Simply monitoring the flow rate of the incoming stream would not be adequate. It is also necessary to know how the disturbance and the manipulative variables affect the process.

2.3.3 Combined Feedforward-Feedback Control

In combined feedforward-feedback control, the manipulative variable is changed by both the feedforward controller and the feedback controller. Most feedforward control systems are installed as combined feedforward-feedback systems (Luyben, 1973). Such a system is shown in block diagram form in Figure 2-4. The Laplace domain variables L(s), M(s), X(s), and $X_{set}(s)$ are the load disturbance, the manipulative variable, the output variable, and the set point, respectively. Transfer functions shown in Figure 2-4 are F(s) for the feedforward controller and B(s) for the feedback controller. The load disturbance is fed to the process through the transfer function $G_L(s)$. The manipulative variable is introduced through the transfer function $G_M(s)$.

2.3.4 Cascade Control

Another type of control frequently used is called cascade control and involves two feedback controllers. In cascade control, the output of one controller (the master controller) is the set point for another controller

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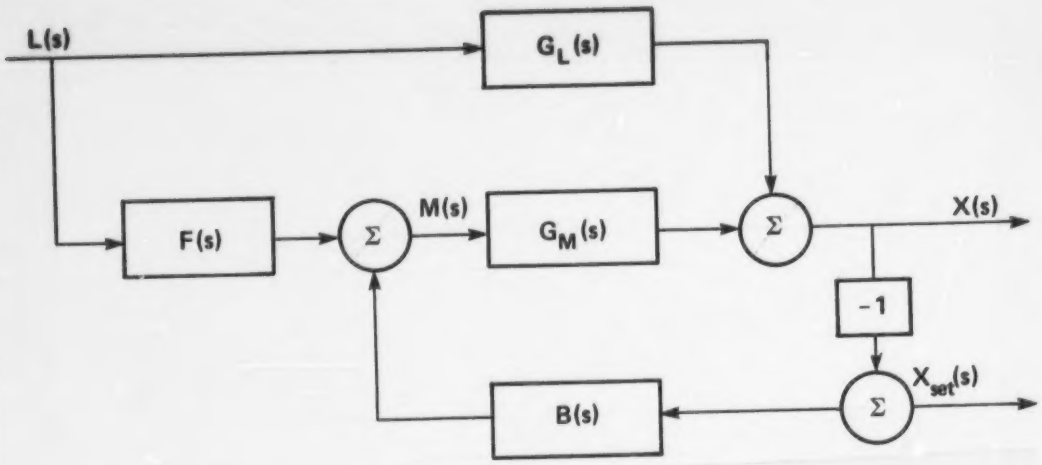

Figure 2-4. Block diagram showing transfer functions for feedforward/feedback control.

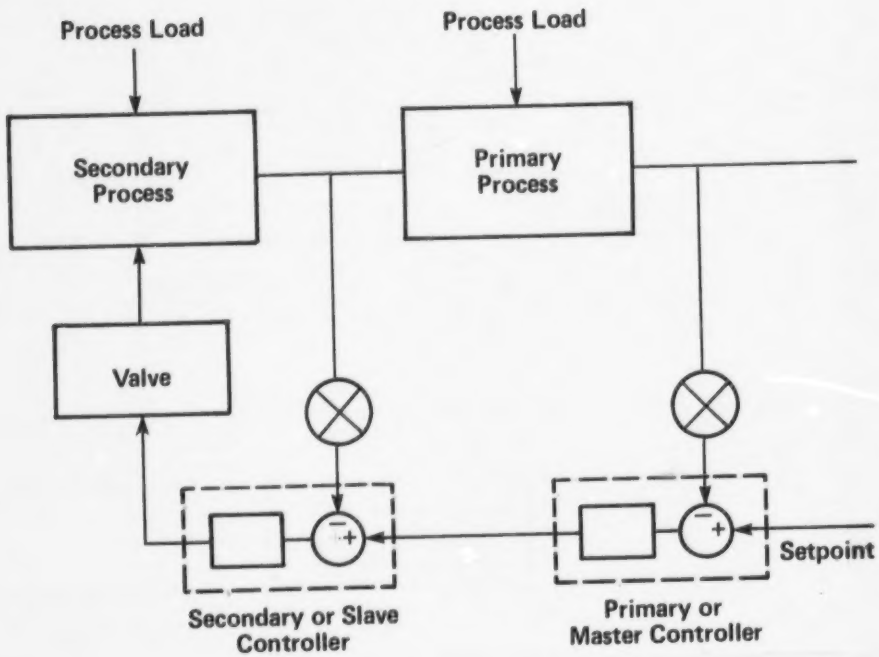

Figure 2-5. Cascade control configuration (from "Guidelines for Control System Design," © reprinted with permission by the publisher, Instrument Society of America, 1972).

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(the slave controller). The primary prpose of cascade control is the elimination of minor disturbances to the process (Luyben, 1973). Cascade control is illustrated in Figure 2-5 (see also discussion by Myron, 1972).

2.4 CONTROL SYSTEM STABILITY

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A process is said to be unstable if its output continues to increase. A linear process is at the limit of stability if it oscillates even when undisturbed, and the amplitude of the oscillation does not decay. In a stable system, the output response is bounded for all bounded inputs. Stability should be considered a constraint on the control systems designer. Both stable and unstable responses are illustrated in Figure 2-6.

A stable system is one which after being perturbed eventually comes to rest. In response to either set point changes or disturbances, the system should neither hunt nor cycle, neither should it be so stable that it is sluggish and does not provide satisfactory control (Buckley, 1964). Most processes are open-loop stable, i.e., stable without controllers. Real systems can be made closed-loop unstable, i.e., unstable with a feedback controller (Luyben, 1973). Conditions of instability may arise in feedback systems due to unfortunate timing between the feedback variable and the input variable. In addition to taking care in the specification of the feedback path, the user of the feedback control must be keenly aware of the strong possibility that even when using only stable components, he may create an unstable system or, as is more common create a system that might become unstable. The stability of the controlled system can be determined by applying a unit impulse to the system in equilibrium and examining the output as time increases. If the perturbation of the output about the equilibrium value approaches zero, the system is stable. If the output increases indefinitely with time, the system is unstable.

Stability may be addressed from a mathematical standpoint. Methods such as the pole zero or root locus provide basic insight into stability. However, less involved approaches are adequate to permit a designer to synthesize a system that has satisfactory short- and long-term behavior consistent with a reasonable degree of stability. A system is stable if its characteristic equation does not have any positive real roots. The

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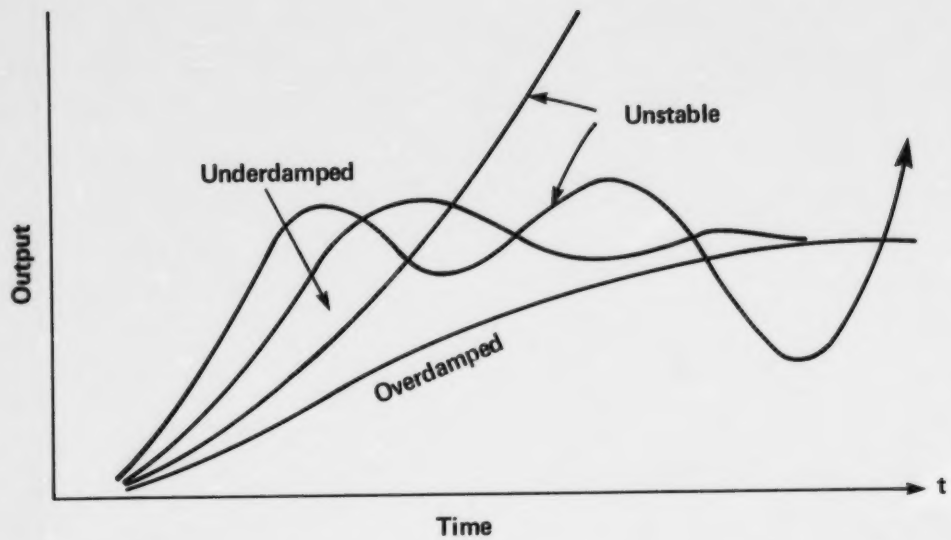

Figure 2-6. Illustrations of outputs from stable and unstable systems (from *Process Modeling, Simulation, and Control for Chemical Engineers*, Luyben, © reprinted with permission by the publishers, McGraw-Hill Company, 1973).

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Routh-Hurwitz tests (discussed in Appendix A-2) provide a check of absolute stability. Relative stability may be characterized using frequency response methods. The Nyquist stability criterion as illustrated in Section 5 may be used to determine relative stability.

2.5 CONTROL SYSTEM PERFORMANCE ANALYSIS

Once a control system is defined through dynamic equations and the necessary transfer functions are determined, the question must be answered, "How well does the system perform the task for which it was designed?"

Since control systems are inherently dynamic systems, the performance is usually specified in terms of both the time response for a specific input signal and the resulting steady-state error. Usually, compromise among these performance parameters is necessary.

2.5.1 Test Input Signals

The system response is usually tested by using standard test input signals, since there is a reasonable correlation between the response of the system to a standard test input and its ability to perform under normal operating conditions. The input signals commonly used are (1) the step input, (2) the ramp input, (3) the parabolic input (the integral of the ramp input), and (4) unit impulse function. These standard input functions, frequently referred to as forcing functions, are illustrated in Figure 2-7. The traditional test input signal for evaluation of chemical process control systems is a step change in set point. The step function can be approximated very closely in practice, as in the sudden opening of a valve.

A number of time domain specifications that are used as performance criteria for closed-loop responses are defined below:

Overshoot: the amount of swing past the set point;

Rise time (speed of response): the time it takes the process to come up to the new set point;

Decay ratio: the ratio of maximum amplitudes of successive oscillations;

Settling time: the time it takes the amplitude of the oscillations to decay to some fraction (0.05) of the change in set point;

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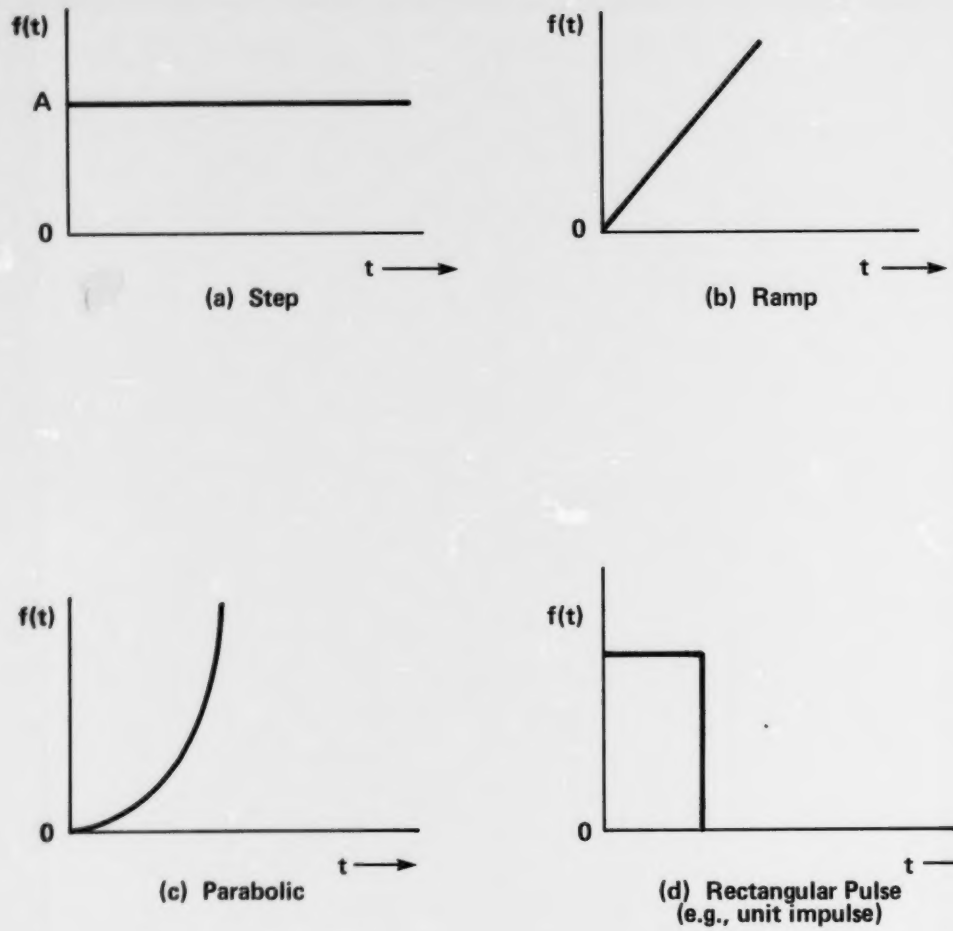

Figure 2-7. Standard input functions.

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ISE: Integral of the squared error.
(Luyben, 1973).

Figure 2-8 illustrates several performance measures defined in terms of the response f(t) of a system to a step input (Dorf, 1967). As shown in the figure, the swiftness of the response is measured by the rise time, T_r, and the peak time period, T_p. For underdamped systems with an overshoot, the 0-100 percent rise time is a useful index. If the system is overdamped, then the peak time is not defined, and the 10-90 percent rise time, T_{r1}, is normally used. The similarity with which the actual response matches the step-input is measured by the percent overshoot and settling time, T_s. The percent overshoot, P.O., is defined as

$$P.O. = \left(\frac{M_p - 1}{1}\right) \times 100\%$$

for a unit step-input, where M_p = the peak value of the time response. The settling time, T_s, is defined as the time required for the system to settle within a certain percentage of the input amplitude (see Figure 2-8).

The steady-state error is apparent for the response depicted in Figure 2-8 as the difference between the desired output response and the actual output response after the settling time.

2.5.2 Performance Index

A performance index is a quantitative measure of the performance of a system (Dorf, 1967). One index of suitable performance is the integral of the square of the error, ISE, which is defined as:

$$ISE = \int_0^T e^2(t)dt \ .$$

The upper limit, T, is a finite time chosen so that the integral approaches a steady-state value. The response, f(t), to a step function, r(t), is shown in Figure 2-9, along with the calculated error, e^2, and ISE responses. It is usually convenient to choose T as the settling time.

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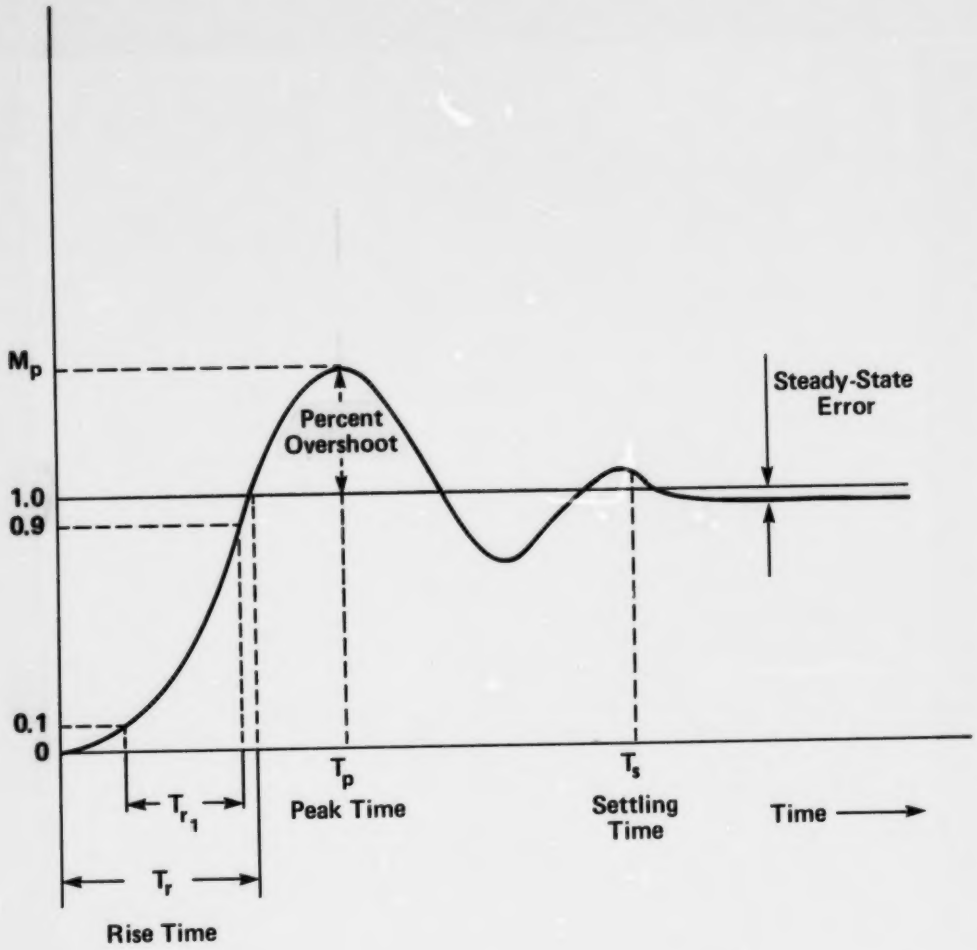

Figure 2-8. Response of a control system to a unit step input

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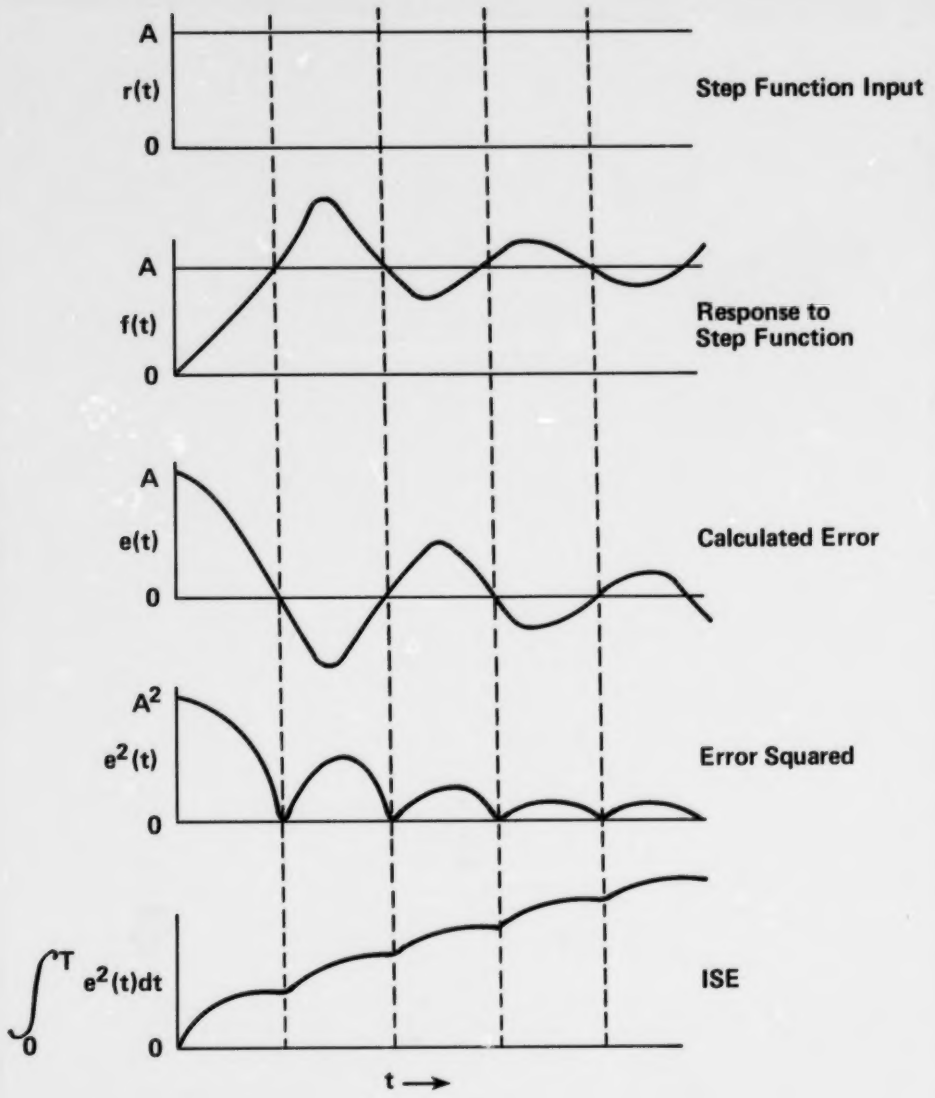

Figure 2-9. The calculation of the integral squared error (ISE) of the response to a step function

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3. CONTROL INSTRUMENTATION AND CONTROLLERS

3.1 THE BASIC CONTROL LOOP

The basic control loop consists of the following elements:

1. A sensor to detect the process variable;
2. A transmitter to convert the sensor signal into an equivalent air pressure or electrical signal;
3. A controller that compares this process signal with a set point and produces an appropriate control signal;
4. A final control element that changes the manipulative variable (Luyben, 1973).

The sensor, transmitter, and control valve, or other final control elements, are physically located on the process equipment. The controller is usually located on a panel in a control room away from the process equipment.

Common sensors include those for flow, stream composition, temperature, pressure, liquid level, pH, and viscosity. For control purposes, it may be necessary to detect only changes in flow rates, temperature, pressure, etc., rather than absolute values of these parameters. The most common type of flow-rate sensor involves orifice plates which allow differential pressure measurements that can be related to flow rates. Venturi tubes and turbine meters are also used. Pressure and differential pressure may be determined using Bourdon tubes, bellows, and diaphragms. Liquid levels are monitored by following the position of a float or by measuring the difference in static pressure between a point in the vapor above the liquid and at an elevation below the liquid surface. For certain process fluids, nuclear radiation gauges are used to detect interfaces and levels.

The transmitter is the interface between the process and the control system. Often called a transducer, the job of the transmitter is to convert the sensor signal into a control signal. The dynamic responses of most sensors and transmitters are usually much faster than the dynamics of the

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process itself. Temperature sensors, typically thermocouples, are a notable exception.

In the controller or comparator, the process signal from the transmitter is compared with the desired value, the set point. If there is any difference between the measured variable and the set point, an error signal is generated and fed to the final control element. Digital real-time microprocessors and minicomputers are being used in increasing numbers to perform the controller function., i.e., to establish the signal to the final controlling element. In fact, the application of "direct digital control" is the most important recent addition to auto-control. The ability to account for many variables simultaneously and to perform the associated complex computations makes direct digital control a powerful and cost-effective tool in auto-control.

Often in industrial process control, the final control element is an air or electrical solenoid-operated control valve that opens and closes to adjust the flow rate. If the variable to be controlled is temperature, the final control element might be a variable transformer that can be used to adjust the current to a resistance heating element.

The number and variety of currently available measurement and control devices is considerable, and the choices are continually expanding. Information is available from a number of sources. The Instrument Society of America (ISA) maintains a Directory of Instrumentation that provides a comprehensive guide to instrumentation products and manufacturers. In addition, the ISA sponsors short courses and annual symposia that focus on instrumentation design and uses in specific areas. Handbooks are available with charts and tables giving control valve performance data. Specific vendors also supply performance specifications for their equipment.

3.2 TYPES OF CONTROLLERS

Three basic types of controllers are frequently used in industrial processes. These types of controllers are distinguished by the types of control action accomplished. The terminology and definitions given below are those established by the ISA (1979).

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1. Proportional Control Action--Control action in which there is a continuous linear relation between the input and output.
2. Derivative (rate) Control Action--In process instrumentation, control action in which the output is proportional to the rate of change of the input.

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3. Integral (reset) Control Action--Control action in which the output is proportional to the time integral of the input; i.e., the rate of change of output is proportional to the input.

The three actions described above may be used individually or combined in three basic commercial controllers (see Coughanowr and Koppel, 1965):

1. Proportional (P): with proportional action; i.e., output signal directly proportional to the error signal.
2. Proportional-integral (PI): with proportional plus integral action.

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3. Proportional-integral-derivative (PID): with all three actions.

The response of these various control systems to a disturbance is illustrated in Figure 3-1. The transfer functions for the various controller types are given in Table 3-1.

3.3 CONTROLLER TUNING

Adjusting the controller settings of an existing process control system to meet a performance criterion is known as controller tuning. The controller settings of major interest are the proportional (or ultimate) gain, the integral time, and the derivative time. Controller tuning is often accomplished empirically. There are, however, some basic rules-of-thumb for various tests of controllers which can be used to establish reasonable settings. One widely used tuning method is the ultimate sensitivity method which is based on the closed loop response of a system to a series of small step changes. The method consists of first finding the ultimate gain, k_u, which is the value of gain where the loop is at the limit of stability with a proportional only feedback controller. The period of the resulting oscillation, p_u, is observed (minutes/cycle). The value of the ultimate gain and the ultimate period are then used to determine the controller settings that will give an approximate quarter amplitude damped response.

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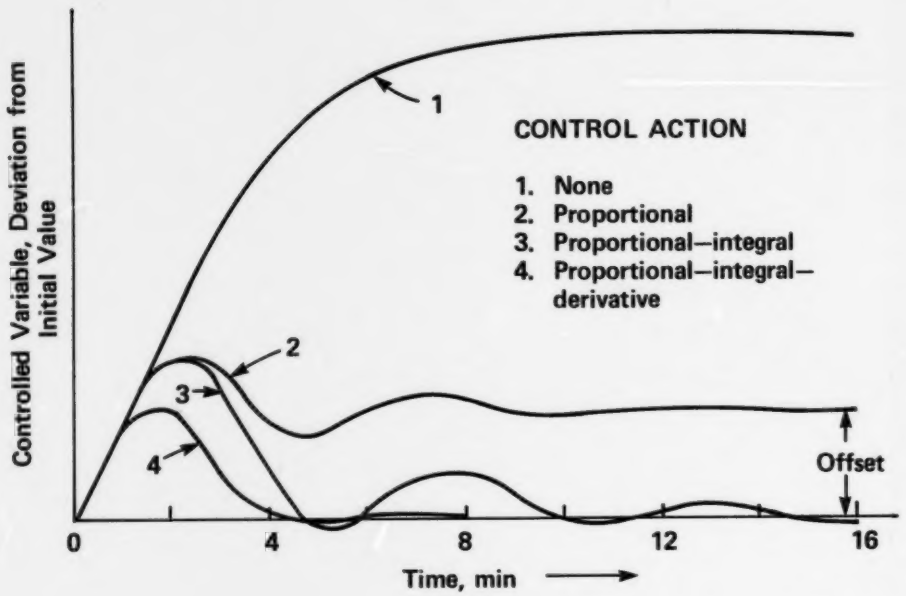

Figure 3-1. Response of the various control systems

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TABLE 3-1. TRANSFER FUNCTIONS FOR CONTROLLERS

Controller type	Transfer function, B(s)[a]
Proportional (P)	$B(s) = k_c$
Integral	$B(s) = \frac{1}{\tau_I s}$
Proportional-Integral (PI)	$B(s) = K_c \frac{\tau_I s + 1}{\tau_I s}$
Derivative (ideal)	$B(s) = \tau_D s$
Derivative (approximate)[b]	$B(s) = \frac{\tau_D s + 1}{\alpha \tau_D s + 1}$
Proportional-Integral-Derivative (PID)	$B(s) = K_c \frac{\tau_I s + 1}{\tau_I s} \quad \frac{\tau_D s + 1}{\alpha \tau_D s + 1}$

[a]Notation in the equations is as follows:

k_c = proportional controller gain

τ_I = integral or reset time

τ_D = derivative time

α is in the range 1/6 to 1/20 for commercial controllers.

[b]This describes an approximate derivative action based on lead-lag units.

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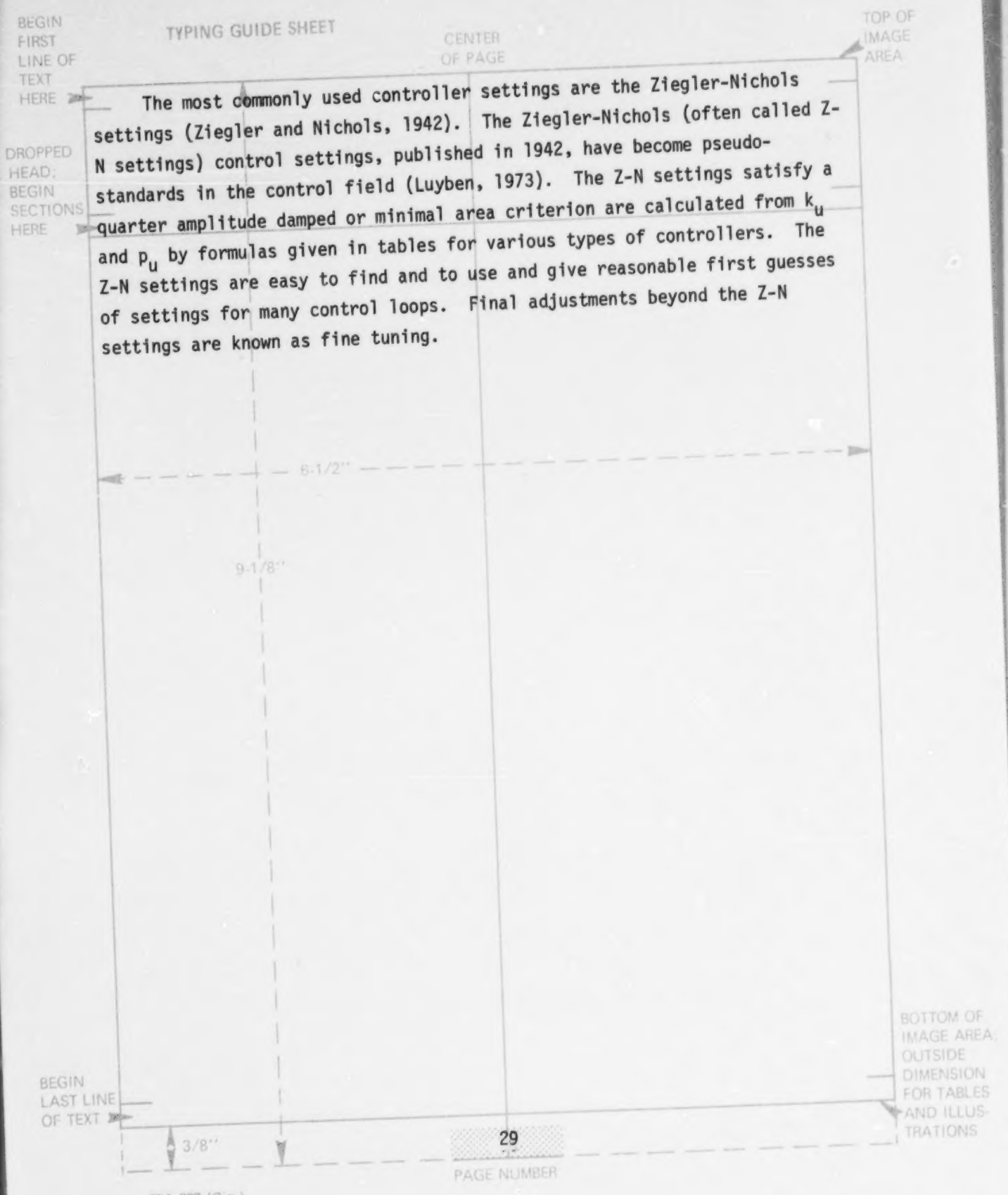
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The most commonly used controller settings are the Ziegler-Nichols settings (Ziegler and Nichols, 1942). The Ziegler-Nichols (often called Z-N settings) control settings, published in 1942, have become pseudo-standards in the control field (Luyben, 1973). The Z-N settings satisfy a quarter amplitude damped or minimal area criterion are calculated from k_u and p_u by formulas given in tables for various types of controllers. The Z-N settings are easy to find and to use and give reasonable first guesses of settings for many control loops. Final adjustments beyond the Z-N settings are known as fine tuning.
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4. DEVELOPMENT OF THE CLOSED-LOOP TRANSFER FUNCTION

CASE STUDY: BLAST FURNACE SCRUBBER WATER RECYCLE SYSTEM

System Description

Iron ore pellets, sinter, fluxing agents, and coke are charged to the blast furnace for the production of iron. The coke is burned to provide heat as well as the reducing environment necessary for the reduction of iron ore (normally in the form of oxides) to molten metallic iron. The molten iron is withdrawn from the bottom of the blast furnace for further processing.

Exhaust gases from the blast furnace are heavily laden with particulate material and are highly contaminated with organic materials as well. These gases are normally heat-exchanged against incoming gas for energy conservation and then scrubbed with an aqueous solution to remove particulate matter and organics. After scrubbing, the gas is discharged to the atmosphere. Pollutant-laden scrubber water is typically processed as shown in Figure 4-1.

A flocculating agent is added to the scrubber effluent to coalesce the fine particles. The effluent, which is then very high in suspended solids, is sent to a thickener. Thickener overflow is directed to a cooling tower. Underflow from the thickener is filtered, and the filtrate is recycled back to the thickener. The wet filter cake from the thickener is a solid discharge from the process.

In the cooling tower, clean effluent is cooled to the temperature appropriate for feed to the furnace gas scrubber. A small makeup water stream (to replace that lost in the cooling tower and filter cake) is added to the cooled effluent, and the combined stream is fed to the scrubber completing the loop.

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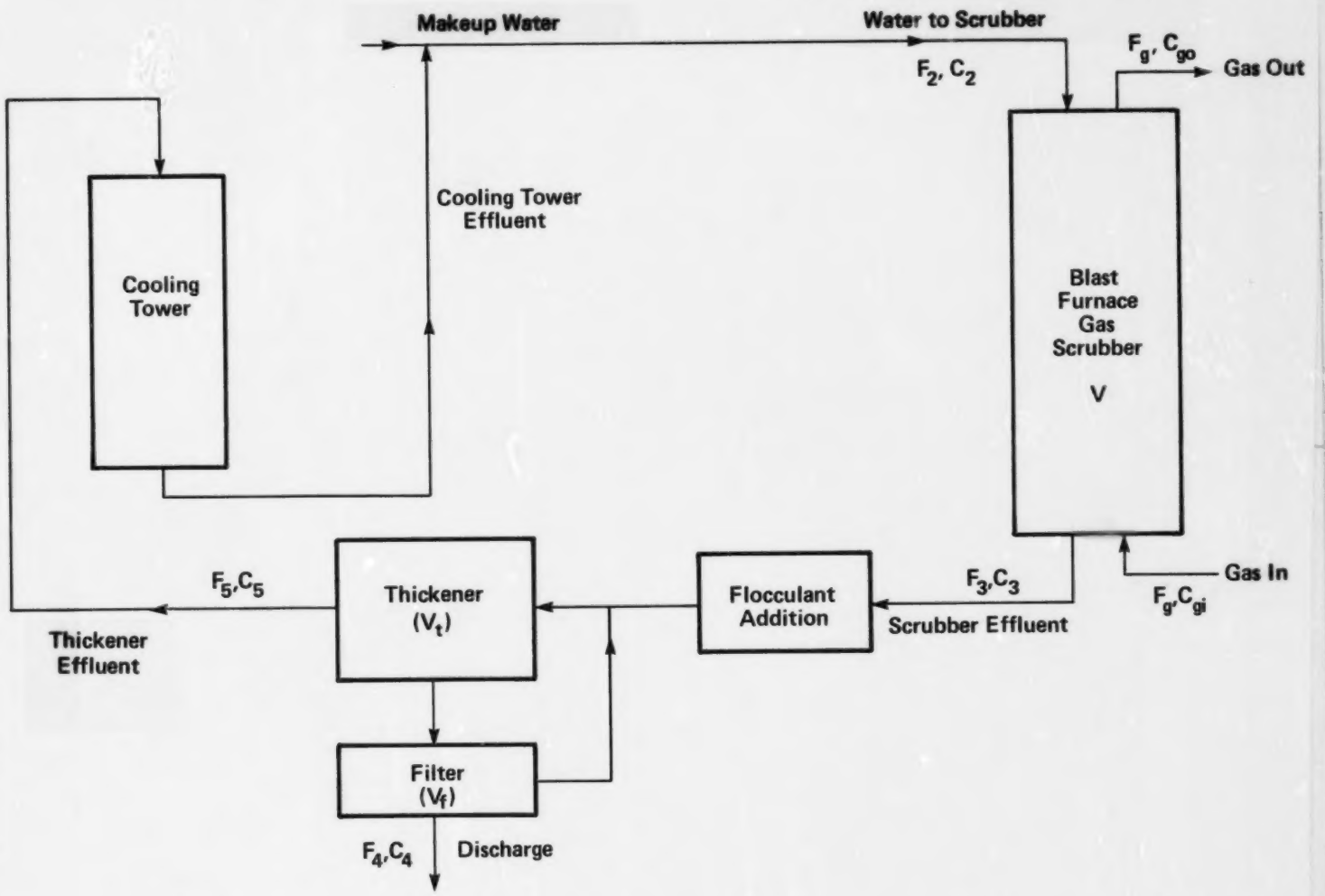

Figure 4-1. Simplified schematic of blast furnace scrubber water recycle loop.

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Dynamic Model and Laplace Transforming

Ultimately, what is desired from the analysis that follows is the closed-loop transfer function for the furnace gas scrubber. This transfer function, G(s), is the ratio of the Laplace transform of the deviation from steady state of the pollutant concentration in the effluent leaving the scrubber to the Laplace transform of the deviation from steady state of C_1, where C_1 is the rate of contaminant removal from the gas divided by the liquid flow rate. This is made clearer by Equations (4-1) and (4-2), and Figure 4-1.

$$G(s) = \frac{C_3(s)}{C_1(s)} \quad . \tag{4-1}$$

Control of the particulate and organic pollutants in the gas leaving the scrubber is accomplished through control of the various parameters that affect the transfer function. The objective is to maintain a steady rate of pollutant removal regardless of excursions in the pollutant loading of the gas entering the scrubber.

The first step in the synthesis of an automated control system is the development of a transient performance model for the uncontrolled process system.

Consider first the wet scrubber. For a simplified analysis, assume that the volume flow rates of both gas and liquid through the absorber are constant.

$$F_g(C_{gi} - C_{go}) = \text{Rate of contaminant removal from gas} = F_1C_1, \tag{4-2}$$

where

C_{gi} = initial pollutant concentration in gas entering the scrubber;
C_{go} = pollutant concentration in gas leaving the scrubber;
F_g = volumetric flow rate of the gas through the scrubber;
F_1 = volumetric flow rate of the liquid through the scrubber.

The term F_1C_1 designates the total quantity of material transferred to the liquid phase in the scrubber. At steady state,

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$$F_1C_1 = F_3C_3 - F_2C_2 \quad \text{or} \quad F_1C_1 = F_1(C_3 - C_2) \qquad (4\text{-}3)$$

where C_3 = pollutant concentration in the aqueous stream leaving the absorber; and

C_2 = pollutant concentration in the aqueous stream entering the absorber.

Variations in the mass loading of the contaminants in the gas, however, will result in perturbations in the quantity of material absorbed by the liquid. These perturbations will be transmitted through the entire recycle loop and consequently will affect pollutant loading in the aqueous feed to the scrubber. The effect, however, will not be apparent in the feed immediately.

The aqueous phase is much more nearly in plug flow than in mixed flow; however, since its residence time is short, it may be acceptable to assume a mixed aqueous phase in the absorber. This simplifies subsequent analysis considerably. So proceeding (referring to Figure 4-1), a transient mass balance on the pollutant loading of the aqueous phase gives:

$$\underbrace{V\frac{dC_3}{dt}}_{\substack{\text{Aqueous phase} \\ \text{accumulation}}} = \underbrace{F_2C_2}_{\substack{\text{Aqueous} \\ \text{input}}} + \underbrace{F_1C_1}_{\substack{\text{Mass transfer} \\ \text{input}}} - \underbrace{F_3C_3}_{\substack{\text{Aqueous} \\ \text{discharge}}}, \qquad (4\text{-}4)$$

where V = aqueous retention volume in the scrubber.

The next step is to Laplace transform Equation (4-4). The resulting transform is given by Equation (4-5).

$$V[sC_3(s) - C_3(o)] = F_2C_2(s) + F_1C_1(s) - F_3C_3(s) . \qquad (4\text{-}5)$$

Hereafter two conventions will be adopted:

1. Ignore the initial conditions in transforming the derivative--this is equivalent to assuming that the response of interest is sufficiently long-term that initial transients have completely decayed.

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2. Designate the transformed variable (in the s-domain) by the same symbol as for the untransformed variable (in the t-domain).

With these conventions and bringing C_3 terms to the left side in Equation (4-5) gives

$$C_3(V_T s + F_3) = F_1C_1 + F_2C_2 \ . \qquad (4\text{-}6)$$

Considering next the streams for flocculant addition, thickener, and filter systems shown in Figure 4-1, a pollutant mass balance on the lumped system gives

$$V_T \frac{dc_{avg}}{dt} = F_3C_3 - F_5C_5 - F_4C_4 \ . \qquad (4\text{-}7)$$

Accumulation in lumped system	Input from scrubber	Effluent from thickener overflow	Effluent discharged with filter cake

The accumulation term in Equation (4-7) requires more development; C_{avg} is really an average concentration for the entire lumped system comprised of the thickener, filter, and filtrate recycle. In Equation (4-7), the volume associated with flocculant addition is neglected since it is very small compared to the other flows. The term V_T is the combined aqueous holdup of the thickener and filter.

$$V_T \, C_{avg} = V_tC_5 + V_fC_4 \qquad (4\text{-}8)$$

where V_f = filter liquid volume and

V_t = thickener liquid volume. In addition, a performance model for the thickener filter combination is needed; for the moment, assume the simplest--a linear relationship between the concentrations of overflow and filtrate.

$$C_4 = kC_5 \ . \qquad (4\text{-}9)$$

Equations (4-8) and (4-9) can now be used to eliminate C_4 and V_{avg} from Equation (4-7), thus giving a single expression relating C_3 and C_5. So

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proceeding, and Laplace transforming the resulting equation gives an expression with C_3 and C_5 as functions of s. With the conventions introduced previously,

$$(V_t + V_f\, k)\, s\, C_5 + F_5\, C_5 = F_3 C_3 - F_4\, k\, C_5 \,. \qquad (4\text{-}10)$$

Or, in transfer function form:

$$\frac{C_5}{C_3} = \frac{F_3}{(V_t + V_f\, k)s + (F_5 + F_4\, k)} \,. \qquad (4\text{-}11)$$

The cooling tower and makeup water stream may be considered together, and

$$C_5 = C_2 \qquad (4\text{-}12)$$

if the following assumptions are made: (1) there is negligible pollutant loss in the cooling tower; and (2) the volume of water discharged with the filter cake can be negelected (i.e., $F_5 \sim F_3$).

From Equation (4-11),

$$\frac{C_2}{C_3} \text{ is of the form } \frac{a_1}{a_2 s + a_3} \,.$$

The concentration, C_2, can now be eliminated utilizing Equations (4-6), (4-11), and (4-12) to give

$$C_3(V_1 s + F_3) = F_1 C_1 + F_2 F_3 \frac{C_3}{(V_t + V_f\, k)s + (F_5 + F_4\, k)} \qquad (4\text{-}13)$$

Simplification will give a closed-loop transfer function (C_3/C_1) for the scrubber system.

$$G(s) = \frac{C_3}{C_1} = \frac{F_1(V_c + V_f)s + F_1(F_5 + F_4\, k)}{V(V_c + V_f)s^2 + V_1 s(F_5 + F_4 + F_3) + F_3(F_5 + F_4\, k)} \qquad (4\text{-}14)$$

This transfer function is oversimplified becasue of the several assumptions introduced in its derivation. More detailed consideration of the performance of each of the process components would lead to a somewhat

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more complicated but more accurate expression for the scrubber transfer function. Nonetheless, the transfer function resulting from the simplified analysis suggests no unusual process transient operation. The denominator of the function is of higher order in s than the numerator, as expected.

Subsequent analysis of the system should involve a stability analysis of the system as indicated, introduction of control systems as appropriate along with their transfer functions, and examination of the stability characteristics of control loops and process loops considered simultaneously.

For example, a very simple control system for the blast furnace scrubber water recycle loop might consist of the following elements:

1. Scrubber sump level controller--a proportional controller for sensing sump liquid level and actuating a control valve to adjust the control valve position, could also be used to control the metering pump discharge for flocculant addition.
2. A cooling tower sump level controller for sensing the level of water in the bottom of the cooling tower and actuating a control valve to influence the cooling tower effluent flow rate. The valve actuator could also be used to control the rate of addition of makeup water to provide the desired flow rate to the scrubber, F_2.

Such a system, though highly oversimplified, represents a reasonable first step in the design of a control system for the scrubber water recycle loop. Each of the two control units has its own unique set of dynamics which can be expressed in terms of its transfer function. In the analysis of the effectiveness of the overall control loop, the controller dynamics are combined with system dynamics in order to assess the overall system response to disturbances. The goal is usually to achieve the dynamic system that responds reasonably quickly to process changes but yet that is not so sensitive as to generate oscillations. Analysis of controller characteristics in conjunction with process characteristics can result in rational selection of controller action and controller settings to achieve the desired system response.

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5. DISTRIBUTED PARAMETER SYSTEM

CASE STUDY: AN ACID GAS REMOVAL SYSTEM

Introduction

This case study is used to illustrate the treatment of a distributed parameter system. Distributed parameter systems are described by partial differential equations in contrast to the ordinary differential equations of a lumped parameter system. If the partial differential equations are linear, or can be linearized, and if the coefficients can be considered to be constant, classical control theory can still be used, although with increased difficulty.

One use of an acid gas removal system might be in conjunction with a coal gasifier. The product of a gasifier normally contains, in addition to the fuel gases, several other gases collectively referred to as the acid gas; typically, these are H_2S, COS, and CO_2. Before the product gases can be processed further, the sulfur-bearing gases must be removed and the CO_2 must also be removed if the product is to be used to form synthetic methane.

The Process

A typical acid gas removal system is the absorber-stripper combination shown schematically in Figure 5-1. The sour (rich) gas from the gasifier enters the absorber at the bottom of the column and contacts a counter-current flow of liquid absorbent (the lean solvent) that enters at the top of the column. The absorbent is chosen for its ability to absorb the specific acid gas(es) of interest. Most of the absorbents are patented and are distributed, along with any specialized operating instructions, by licensed vendors; for example, Rectisol uses cold methanol and Benfield uses hot carbonate. The acid gas is often referred to as the solute.

The "sweet" (lean) gas leaves at the top of the column, and the solvent plus the absorbed gas (the rich solvent) leaves at the bottom. The

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Figure 5-1. Simplified schematic of acid gas removal system.

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rich solvent is then passed through a heater into the top of the stripper column, where it contacts a countercurrent flow of an inert gas, typically nitrogen, that strips the acid gas from the solvent. The stripper is essentially an absorber operated in reverse. The stripped acid gas is removed for further processing and/or disposal, and the lean solvent is recycled through a chiller to the top of the absorber column.

System Model

A simple packed column for continuous contacting of a gas and a liquid is shown in Figure 5-2. For simplicity, a single acid gas is considered in the analysis.

In Figure 5-2,

V' = the gas flow rate,

Y = the concentration of the acid gas in the rich gas,

F' = the liquid (absorbent) flow rate,

X = the concentration of the absorbed acid gas in the absorbent.

A mass balance on the acid gas at a differential slice through the column yields

$$\frac{\partial}{\partial t}(1-\varepsilon)X = -F'\frac{\partial X}{\partial z} + V'\frac{\partial Y}{\partial z} \qquad (5\text{-}1)$$

where ε is the void fraction of the packing, and $(1-\varepsilon)$ represents the holdup in the differential slice.

The two concentrations, X and Y, are functions not only of time but also of the location within the column. If X and Y are expressed as the sum of a steady-state value that is independent of time and time-varying perturbation with respect to that steady-state value, they can be written as

$$X(z,t) = X_s(z) + x(z,t) \qquad (5\text{-}2a)$$

and

$$Y(x,t) = Y_s(z) + y(z,t). \qquad (5\text{-}2b)$$

Substituting in Equation (5-1) yields

$$\frac{\partial}{\partial t}(1-\varepsilon)\,x(z,t) = -F'\frac{dx_s}{dz} - F'\frac{\partial x(z,t)}{\partial z} + V'\frac{dY_s}{dz} + V'\frac{\partial y(z,t)}{\partial z} \qquad (5\text{-}3)$$

The steady-state equation is given below:

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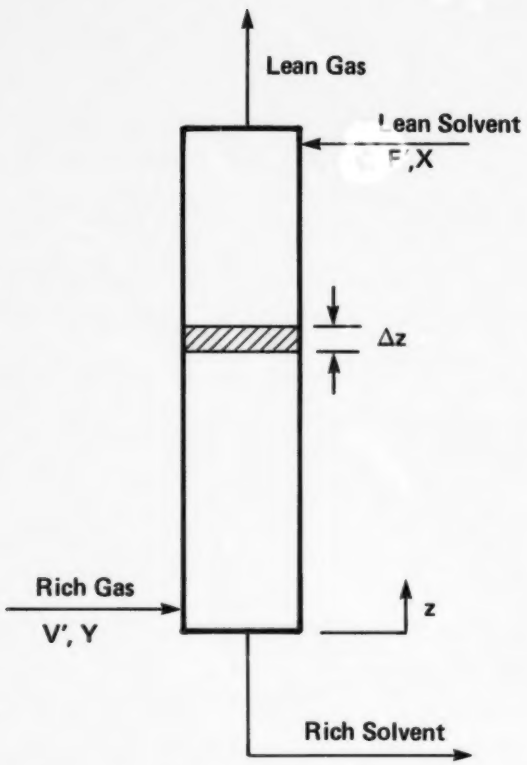

Figure 5-2. The packed-column absorber.

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$$0 = F' \frac{dX_s}{dz} + F' \frac{dY_s}{dz} . \qquad (5\text{-}4)$$

Subtracting Equation (5-4) from Equation (5-3) leaves the dynamic equation:

$$\frac{\partial}{\partial t} (1 - \varepsilon)\, x(z,t) = - F' \frac{\partial x(z,t)}{\partial t} + V' \frac{\partial y(z,t)}{\partial z} . \qquad (5\text{-}5)$$

By assuming a linear equilibrium relationship between the acid gas in the liquid and gas phases,

$$y(z,t) = mx(z,t) \qquad (5\text{-}6)$$

Using Equation (5-6), the dynamic equation becomes

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$$\frac{\partial}{\partial t} (Y/m) = - F \frac{\partial (y/m)}{\partial z} + V \frac{\partial y}{\partial z} \qquad (5\text{-}7)$$

where $F = F'/(1 - \varepsilon)$ and $V = V'/(1 - \varepsilon)$.

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If the slope of the equilibrium curve, m, can be assumed constant, both in time and space, the dynamic equation of the absorber can now be written in its final form as

$$\frac{\partial y(z,t)}{\partial t} = - a \frac{\partial y(z,t)}{\partial z} \qquad (5\text{-}8)$$

where $a = (F - mV)$. Equation (5-8) is in the form of the first-order diffusion equation.

System Analysis

With the further assumption of constant flow rates, the Laplace transformation can be used to obtain a transfer function from Equation (5-8). The transformed equation is:

$$\frac{d\bar{y}(z,s)}{dz} + \frac{s}{a} \bar{y}(z,s) = 0 \qquad (5\text{-}9)$$

This is a first-order homogeneous differential equation with the solution

$$\bar{y}(z,s) = C \exp\left(\frac{-sz}{a}\right) . \qquad (5\text{-}10)$$

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At z = 0, the entry point of the rich gas, $\bar{y}(0,s) = C$, so that

$$\frac{\bar{y}(z,s)}{\bar{y}(0,s)} = \exp\left(-\frac{sz}{a}\right). \qquad (5\text{-}11)$$

Equation (5-11) is a transfer function relating the change in the concentration of the acid gas in the rich gas at <u>any</u> location in the absorber to a change in the concentration of the acid gas <u>entering</u> the absorber. With z set equal to the height H of the absorber, the transfer function relating the change in the exit concentration to a change in the entry concentration becomes

$$\frac{\bar{y}(H,s)}{\bar{y}(0,s)} = \exp\left(-\frac{H}{a}s\right). \qquad (5\text{-}12)$$

Equation (5-12) is the transfer function of a transportation lag with a dead time T equal to H/a. Consequently, any change in the entry concentration at t = 0 would appear at the exit at H/a units of time later without any change in value or modification of shape. A spike, or impulse, representing a 10 percent increase in the input concentration would result in a 10 percent spike in the output concentration in H/a units of time later. Step and sinusoidal changes in the input concentration would be similarly delayed and unchanged.

<u>Stability Analysis</u>

The absorber standing alone may be considered to be stable. However, it has a destabilizing influence on the rest of the system and on any attempts to control the effects of any disturbance or upsets in the input gas concentration. For discussion purposes, the absorber is treated independent of the rest of the system in order to examine its effect on a simplified and hypothetical control system.

The first step in designing a control system, is to develop a control philosophy or strategy based on the performance requirements. Such a strategy includes a determination of the variable to be manipulated; in this case, the absorbent (liquid) flow rate will be used. It will be assumed that only proportional control will be applied, and that all

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other component transfer functions can be represented by simple, first-order time lags with all gains incorporated into a single loop gain, A. The hypothetical regulator is shown in Figure 5-3.

The closed-loop transfer function relating the change in the output concentration to a change in the input concentration can be written as

$$G_R(s) = \frac{Y}{Y_0} = \frac{e^{-Ts}}{1 + \frac{Ae^{-Ts}}{(\tau_1 s+1)(\tau_2 s+1)(\tau_3 s+1)}} \qquad (5\text{-}13)$$

Without the transportation lag (i.e., T = 0), this would be a minimum phase transfer function of a simple third-order system that is conditionally stable. An increase in the value of the gain A would decrease the value of the output gas concentration, but it would also decrease the stability of the system. The transient response would change from that of an overdamped system (no overshoot) to an underdamped (damped oscillations) system and ultimately to an unstable (diverging oscillations) system.

As the time lag, T, becomes finite, the system stability for a stable value of the gain A would decrease; increasing T, without changing A, would eventually drive the system unstable. This can be best seen in the Nyquist diagram of Figure 5-4, which uses the characteristic equation of Equation (5-13). In Figure 5-4a with T = 0, (no transportation lag) and A = A_1, the polar plot does not encircle the -1 and the system is stable. If the gain is increased to A_2, the polar plot still does not encircle the -1 point but does come closer to it, indicating a decrease in the stability of the system (increased oscillations but still damped).

A finite T changes the system from minimum to nonmimimum phase. Figure 5-4b shows the effect of an increasing T upon the stability of the system with the gain held constant at A_1. As T goes from 0 to T_1, the polar plot moves closer to -1 decreasing the system stability. For T = T_2, the plot passes through -1 (undamped oscillations and marginal stability); for T = T_3, it encircles -1 (diverging oscillations and instability).

A root locus sketch of the hypothetical system with T = 0 is shown in Figure 5-5a for an increasing value of the gain A along with the roots for

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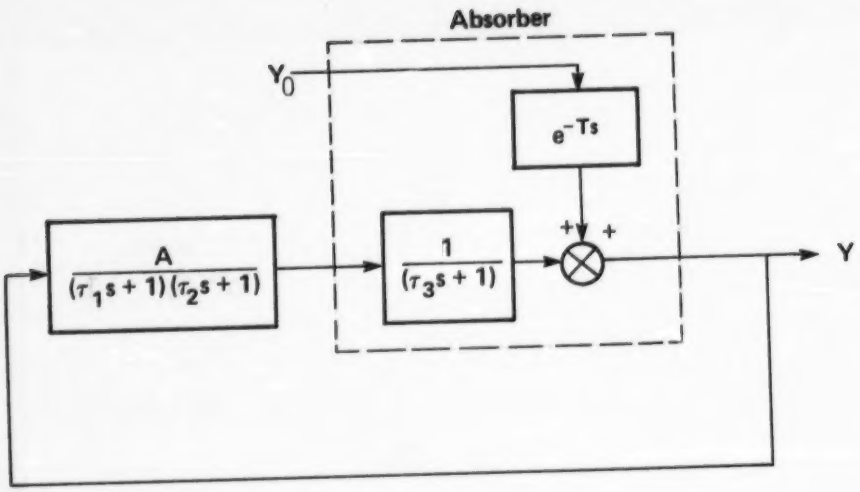

Figure 5-3. A hypothetical absorber control system.

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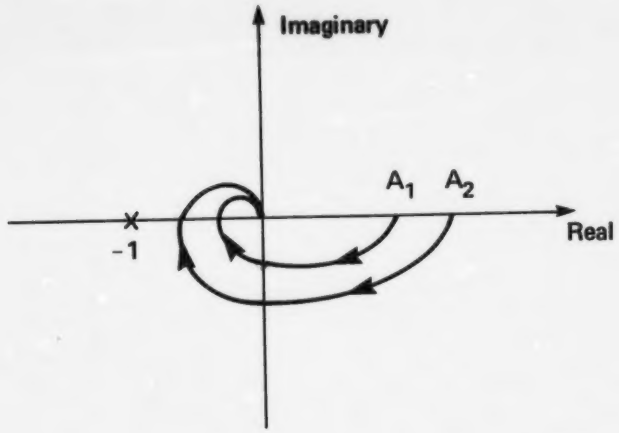

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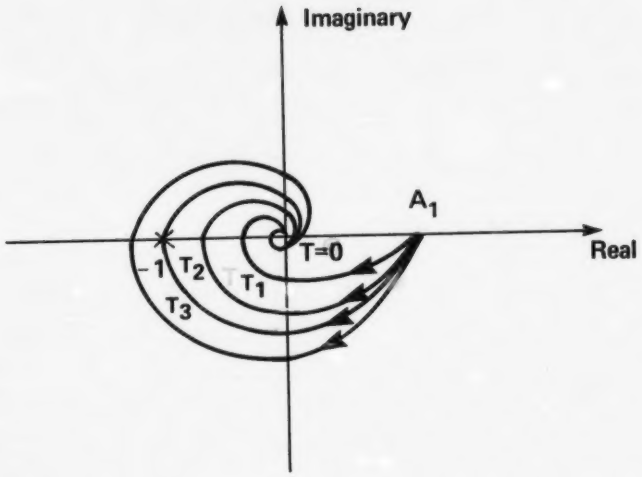

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Figure 5-4. Nyquist diagrams: (a) T = 0; (b) constant gain.

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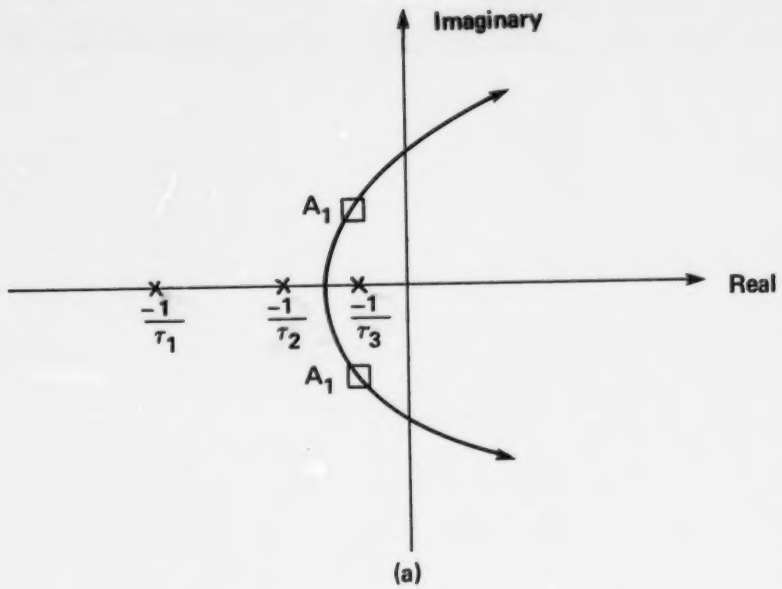

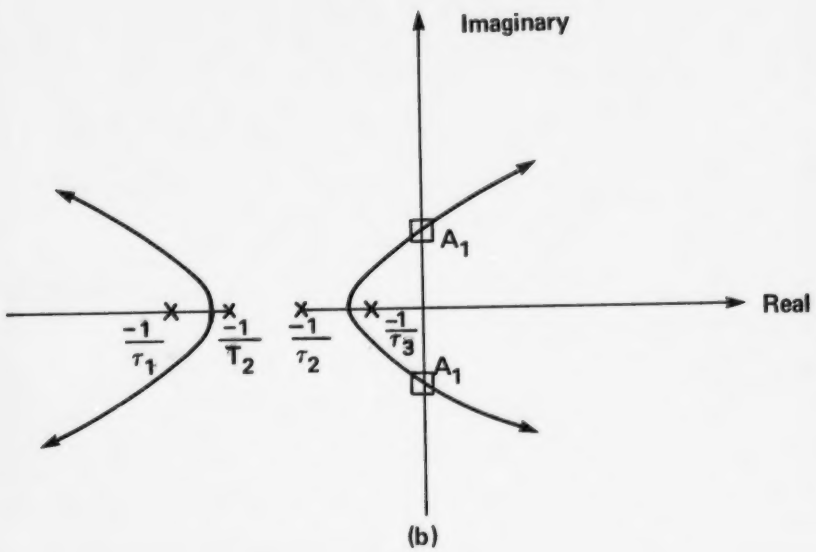

Figure 5-5. Root locus sketches: (a) T = 0; (b) T = T_2.

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a gain value of A_1. A true root locus involving e^{-Ts} has an infinite number of branches and is quite difficult to construct. Consequently, it is customary to approximate the exponential by a rational expression. Of the several choices, the simplest is the single first-order lag approximation

$$e^{-Ts} \simeq \frac{1}{Ts + 1} . \qquad (5\text{-}14)$$

With this approximation and a finite T, a first-order lag is added which translates into the addition of another pole (a destabilizing influence) and the addition of another branch to the root locus. As T increases, this pole approaches the imaginary axis, further destabilizing the system. Figure 5-5b shows that for T_2 the dead time pole has moved the dominant pair of the roots for A_1 to the imaginary axis, a condition of undamped oscillations and marginal stability. This is analogous to the polar plot passing through -1 in the Nyquist diagram of Figure 5-4b. If T were to be further increased, the dominant pole pair for A_1 would move to the right of the imaginary axis and the system would be unstable.

Discussion

Although the model that was used in this study is very simple and represents many assumptions and approximations, it does illustrate the major features of a distributed parameter process and its control. The process is described by partial differential equation(s) that represent a nonminimum phase system, and a time delay or lag that has a strong destabilizing influence on any system of which it is a part.

The next step in complexity in the modeling of an absorber (or stripper) would be the inclusion of mass transfer diffusion with axial dispersion within a single phase. The influence of temperature changes (arising from the heat of absorption) on the equilibrium could also be modeled as a function of acid gas concentration. The inclusion of the interfacial mass transfer between the liquid and gas further increases the complexity of the process model. Nonlinear partial differential equations can be handled by linearization across the entire absorber or by dividing the absorber into two or more sections and then writing the partial

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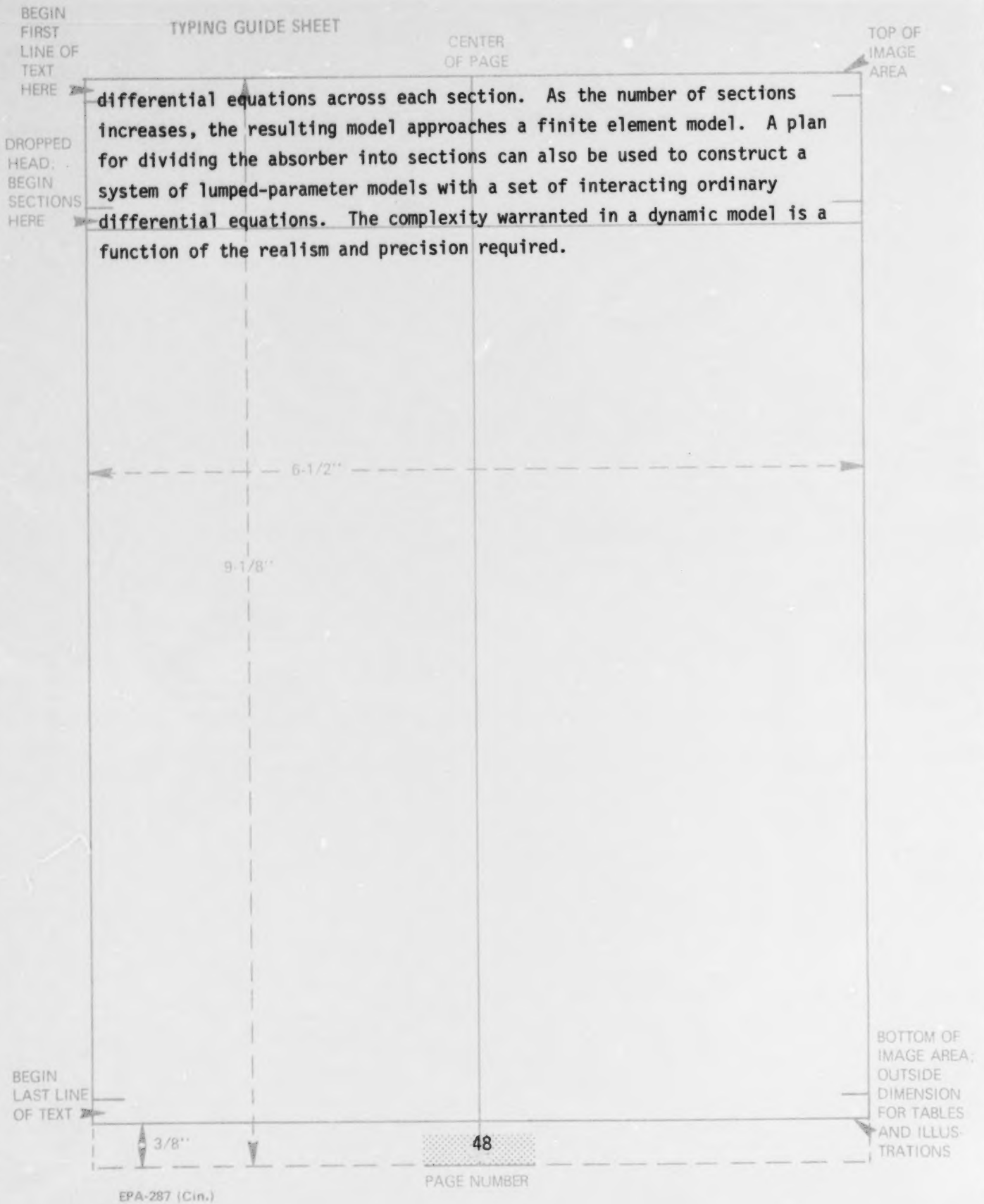
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differential equations across each section. As the number of sections increases, the resulting model approaches a finite element model. A plan for dividing the absorber into sections can also be used to construct a system of lumped-parameter models with a set of interacting ordinary differential equations. The complexity warranted in a dynamic model is a function of the realism and precision required.
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6. ANALYSIS OF FEEDFORWARD CONTROL SYSTEM

CASE STUDY: FLUIDIZED-BED COMBUSTION WITH LIMESTONE CONTROL

System Description

Based on data from pilot and demonstration scale facilities in the United States, fluidized-bed combustion (FBC) appears to be an efficient and economically attractive method for producing useful energy from coal in an environmentally acceptable manner. FBC is used primarily for steam generation. The fluidized bed of the combustor is made up primarily of particles of an inert material (e.g., sand) and/or a sulfur dioxide sorbent such as limestone or dolomite. The crushed coal is present in a smaller amount. The bed material is fluidized by air passing through it. Combustion of the fuel is accomplished at low temperatures, and the resulting sulfur dioxide is effectively adsorbed by the limestone in the bed. Due to the low combustion temperatures, nitrogen oxides in the flue gas are minimal. Corrosion in the combustor is minimized as a result of the low temperatures and low concentrations of sulfur dioxide and nitrogen oxides.

A schematic diagram showing an FBC reactor and feed configuration is given as Figure 6-1. Mass flow rates of coal and limestone are expressed as $\dot{m}_C$ and $\dot{m}_L$, respectively. A nuclear gauge sensor to allow automatic control of the sulfur/limestone ratio in the feed mixture is also indicated on the diagram. The sensor shown here is a continuous online sulfur analyzer utilizing neutron-indiced gamma spectrometry with scintillation counting. The sulfur content of the coal, as indicated by the sensor, is multiplied by the coal flow rate and then compared to the stoichiometric mix required for makeup limestone to insure acceptable adsorption of the sulfur in the fluidized bed to meet environmental standards for flue gas emissions. The approach to automatic control that is illustrated by Figure 6-1 is one of feedforward control.

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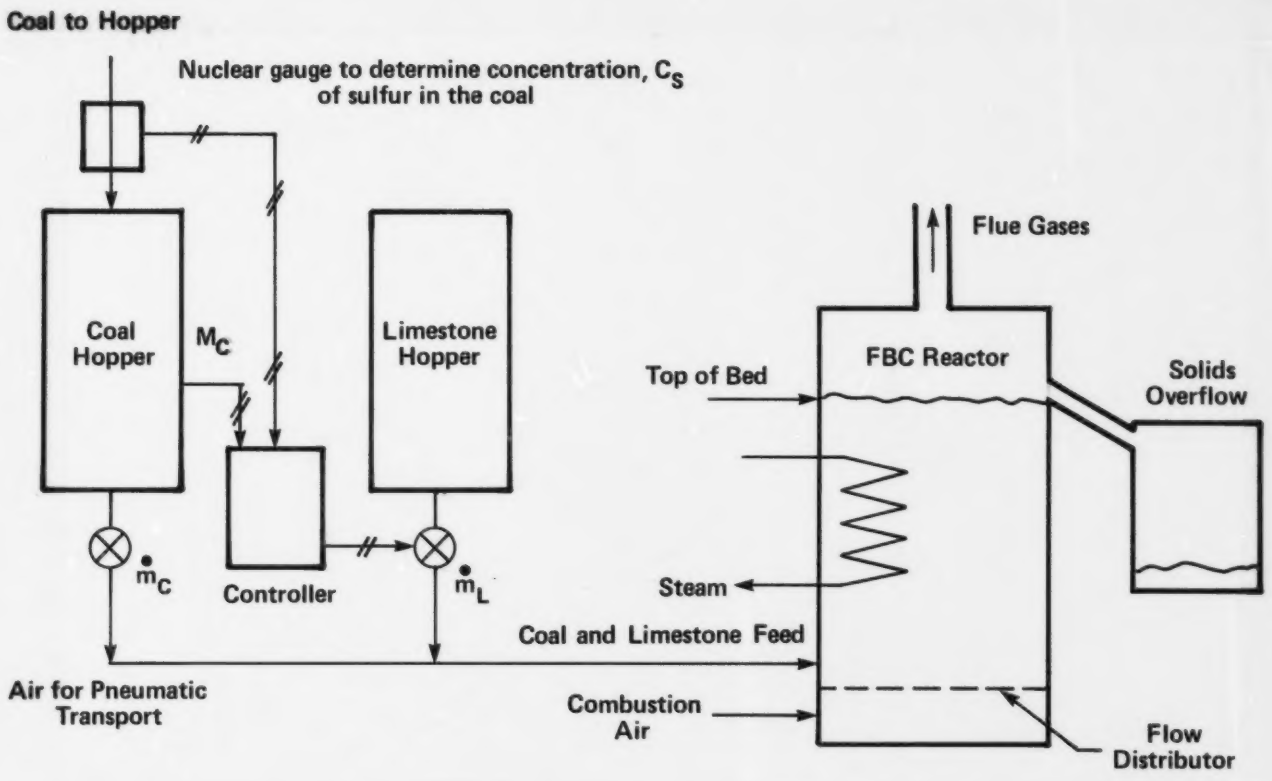

Figure 6-1. Schematic of fluidized-bed combustor.

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Feedforward control is commonly used for throughput changes whereas feedforward control in response to a composition disturbance is only occasionally used because instruments designed for continuous solid composition analysis have been developed only recently.

A more conventional method for controlling the stoichiometric ratio of limestone to sulfur is accomplished via measurement of sulfur dioxide in the exit flue gas with feedback control to the limestone flow valve. However, for very large plants, where coal types (and especially sulfur content of the coals) may fluctuate often, measurement of SO_2 in flue gases means an inherent lag between the time that excess sulfur coal passes the limestone mixing point and the time that the excess sulfur is detected.

The advantage of feedforward control over feedback control is that perfect control can, in theory, be achieved. A disturbance can produce no error in the controlled output variable if the feedforward control is perfect. The disadvantage of feedforward control, besides difficulty in detecting the disturbance, is that the influence of the disturbance on the manipulative variables in the process must be well understood. The transfer functions must be known, at least approximately.

A block diagram of the FBC control system is shown in Figure 6-2. In the block diagram, the representations are:

$M_C(s)$ = Coal flow rate into and out of hopper and to FBC, (tons/hr)(a possible secondary disturbance).

$M_L(s)$ = Limestone (adsorbent) flow rate (tons/hr)--the manipulative variable.

$C_S(s)$ = Sulfur concentration in coal, percent weight. The disturbance to which system responds.

$\begin{cases} G_{SC}(s), \\ G_{LC}(s) \end{cases}$ = Process transfer functions.

$F(s)$ = Transfer function of the feedforward control device.

$X(s)$ = The desired response, or output, of the system.

Henceforth, Laplace domain variables will be understood and written without reference to s.

In this case,

$$X = M_L/C_S M_C - SET_{L/C} \qquad (6\text{-}1)$$

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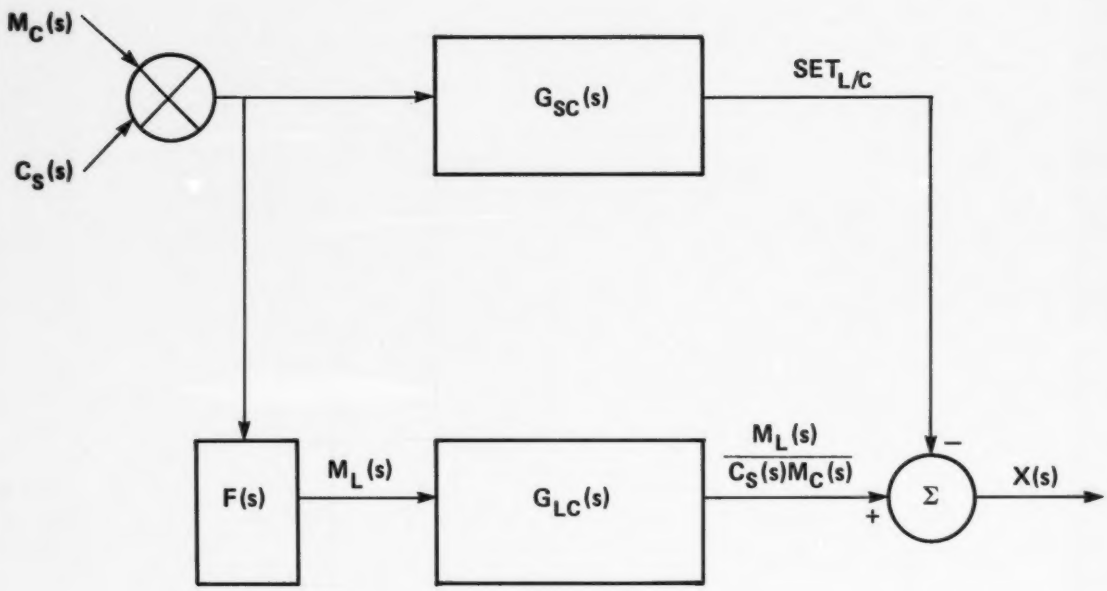

Figure 6-2. Block diagram for the FBC limestone control system.

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where $SET_{L/C}$ = the required stoichiometric ratio of makeup limestone to coal sulfur input to achieve satisfactory sulfur removal. Therefore, for perfect control, X(s) = 0.

Note also that:

$$X = G_{SC} C_S M_C + G_{LC} M_L \quad . \tag{6-2}$$

To obtain the feedforward controller transfer function, F(s), both G_{SC} and G_{LC} must be known. This is expressed, setting X(s) to zero, as

$$F = \frac{M_L}{C_S M_C} = \frac{\text{manipulative variable}}{\text{disturbance}}$$

$$= - \frac{G_{SC}}{G_{LC}} \quad . \tag{6-3}$$

It is seen that

$$G_{LC} = (C_S M_C)^{-1} \quad \text{and} \quad G_{SC} = \frac{SET_{L/C}}{C_S M_C} \tag{6-4}$$

so that the simple transfer function is:

$$F(s) = SET_{L/C} \quad \text{and} \quad G_{LC} = F/M_L, \; G_{SC} = SET_{L/C} \cdot F/M_L. \tag{6-5}$$

Finally,

$$M_L = C_S M_C \; SET_{L/C} \tag{6-6}$$

and the limestone input to the reactor may be controlled within limits of error imposed by measurement, knowledge of the proper solids mix, and system response.

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7. EVALUATION OF CONTROL OPTIONS

CASE STUDY: LIMESTONE SCRUBBER SLURRY

Introduction

This case study is based on results of a research effort performed by Dr. Patrick Garrett of the University of Cincinnati for the U.S. Environmental Protection Agency/IERL-RTP. The research involved the investigation of control options for slurry limestone addition, slurry density, absorber liquid-to-gas ratio, and slurry buffer addition systems (Garrett, 1981). This case study focuses on work related to the slurry limestone addition.

Three control approaches were analyzed and specified for dense limestone addition to the recirculating slurry: (1) stoichiometric control, (2) pH control, and (3) stoichiometric-assisted pH control. These approaches were modeled, simulated using analog computer, and tested at the Shawnee TCA limestone scrubber.

Results of the study indicate that stoichiometric control alone is compromised because internal factors affecting the slurry are not taken into account. It does, however, provide rapid control response to analytic changes. The pH control of limestone feedrate is based on slurry pH measurements and thus takes into account the slurry reaction. A rigorous design, however, is required for stable and responsive performance since the process intervenes between the controller measurement and the limestone feed. Control based on pH is recommended for conditions of high reaction gain (Δ pH/Δ limestone feedrate).

Stoichiometric-assisted pH control was determined to be the optimum control system resulting in maximum reduction of slurry disorder. The increased system complexity associated with this system, however, is justified only under conditions of low slurry reaction gain.

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System Model

Simultaneous slurry dissolution and neutralization reactions and recirculation deadtime complicate the mathematical modeling of the recirculating limestone slurry. Nevertheless, a mathematical model was formulated and subsequently validated. Coefficients were determined using experimental data from the Shawnee TCA limestone scrubber. Transfer functions for dissolution and neutralization were obtained as well as the reaction gain. It was observed that the reaction gain, K_p, decreases as pH is raised. The transfer functions are given in Table 7-1. The first-order time constant, t_d, is the neutralization tank volume, V, divided by the flow rate, F. A block diagram depicting the system is shown in Figure 7-1. An interesting observation is a resonance phenomenon in the limestone dissolution behavior. Operation of the scrubber with load disturbances at about 1-hour intervals can result in an oscillatory response.

Analog Simulation

Four control options for limestone additions were studied using analog computer simulation. The results are shown in Figure 7-2. The control systems were simulated with step changes in load of ± 0.2 pH unit. The slurry pH set point was maintained at 5.5. A steady-state scrubber load of 0.5 pH was maintained. The shortest time for recovery to the operating set point following load disturbance was achieved by the combination feedforward/feedback option. This reflects the ability of feedforward stoichiometric control to respond to inlet changes in sulfur dioxide prior to influencing the slurry. The pH feedback control is geared to solve for slurry unknowns such as fly ash alkali content.

Straight pH feedback was determined to be capable of accurate and well-behaved control, and the controller antiwindup batching feature offers an advantage over conventional proportional plus integral (PI) control. The antiwindup feature disconnects the error signal being sent to the integrator for output activation commands of <0% >100%. Although load increases require approximately the same recovery time with and without the antiwindup feature, recovery from load decreases was shortened notably for the PI with antiwindup.

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TABLE 7-1. TRANSFER FUNCTIONS FOR LIMESTONE SLURRY

Process element	Transfer function	Approximation
Dissolution $G_C(s)$	$\frac{k_1}{1-k_2 \exp(-t_d S)}$	$\frac{(0.037)-(0.037)(0.063)^n}{1 - 0.963}$
Reaction gain K_p	$\frac{\Delta pH}{\Delta feedrate}$	$0.18 \left(\frac{pH}{liters/min}\right)$
Neutralization $G_N(s)$	$\frac{1/t_d S + 1}{1 - 1/t_d S + 1}$	$\frac{1}{10S}$ (min^{-1})

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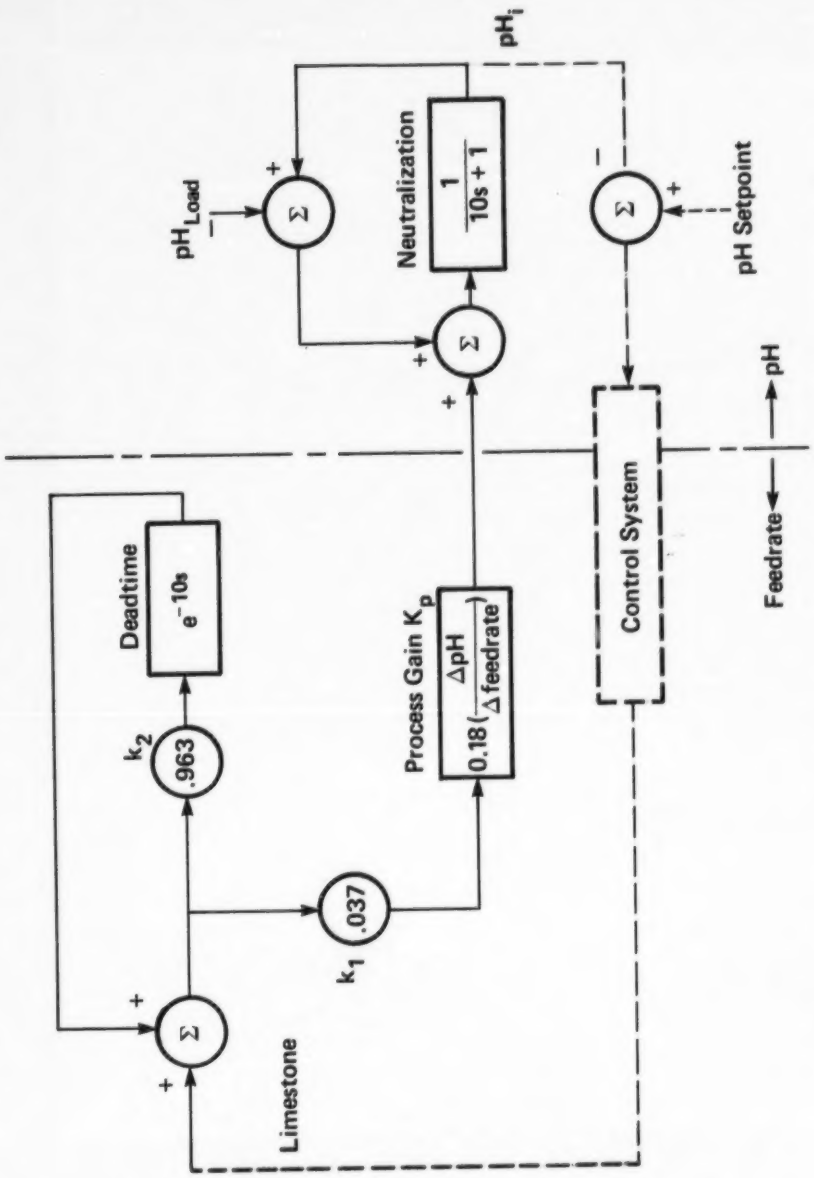

Figure 7-1. Block diagram showing the slurry dissolution and neutralization system (reproduced from Garrett, 1981).

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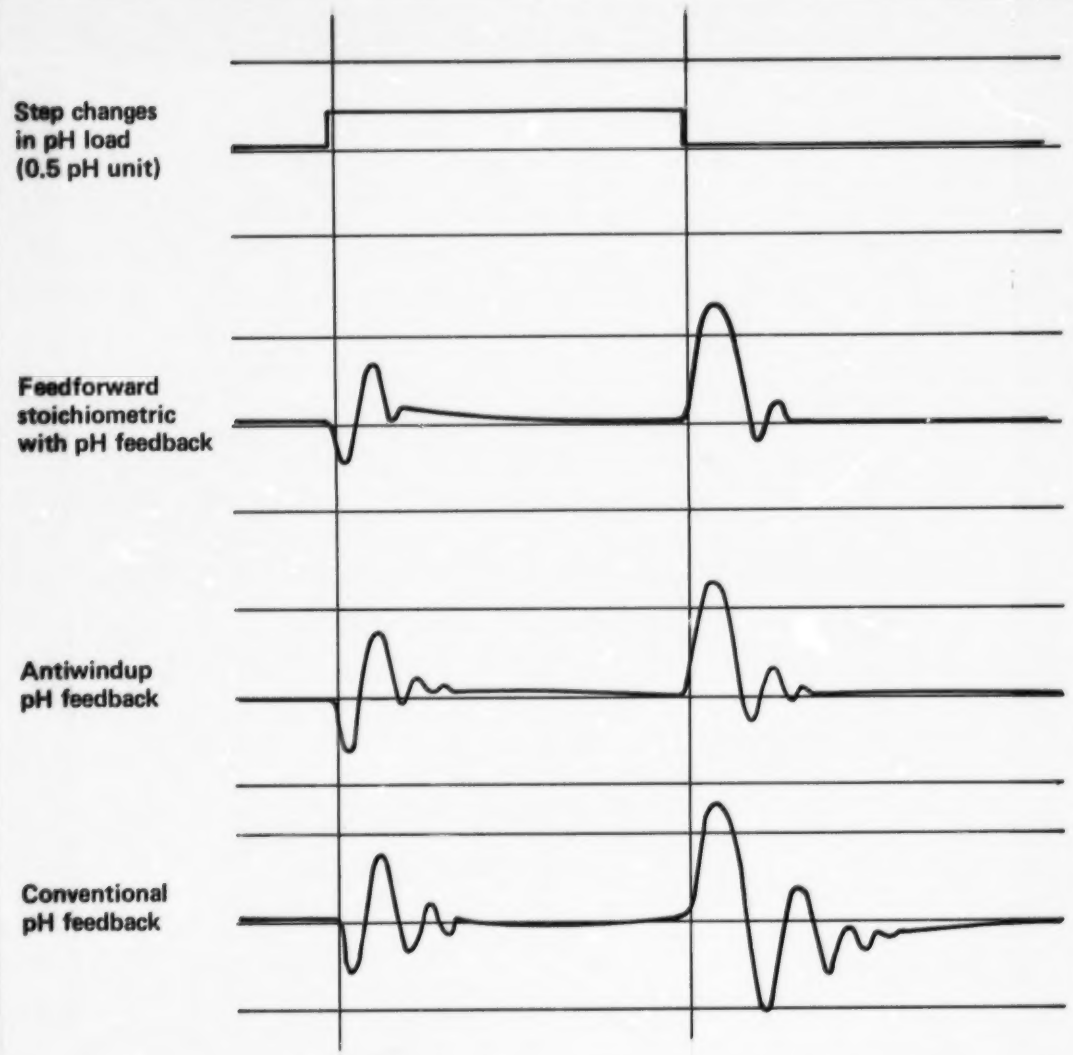
Step changes in pH load (0.5 pH unit)
Feedforward stoichiometric with pH feedback
Antiwindup pH feedback
Conventional pH feedback

Legend: 100 minutes/horizontal division
0.1 pH/vertical division

Figure 7-2. Results of analog computer simulation runs for three control methods; initial pH = 5.5 with load = 0.5 pH unit imposed. (Reproduced from Garrett, 1981).

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Feedforward Stoichiometric Control

Feedforward stoichiometric control for limestone addition is based upon the stoichiometric ratio of limestone to sulfur conrresponding to the desired level of sulfur removal. Limestone is added at a rate proportional to the inlet SO_2 rate. The set point is the desired stoichiometric ratio. This ratio will deviate from the true slurry stoichiometry due to variations in fly ash alkaline content, limestone quality and solubility, and chloride content of the slurry. Feedforward control is useful, however, because of its ability to correct for SO_2 load disturbances entering the recirculating slurry.

Control System Based on pH

Feedback control of the limestone addition by means of a slurry pH measurement can accommodate the following scrubber factors:

Reaction gain,
Fly ash alkalinity,
Calcium sulfite,
Limestone variation,
Slurry chloride content.

A detailed analysis of the scrubber process was performed in order to determine a controller capable of maintaining a required pH set point with unconditional stability.

A proportional plus integral controller is appropriate for feedback control. A narrow range pH probe (4.5 to 6.5) is recommended to increase the sensitivity and control responsiveness to the pH measurement. It is assumed that the pH measurement is an indicator of aggregate slurry chemical reactions. Addition of organic buffer to the slurry in proportion to one-half percent of the limestone feedrate was shown to improve the SO_2 removal to the 90 percent range. The simulated pH feedback slurry control system with buffer is depicted in Figure 7-3.

Stoichiometric-Assisted pH Control

Improved control of the limestone feedrate can be obtained using stoichiometric-assisted pH control. Application of this method is

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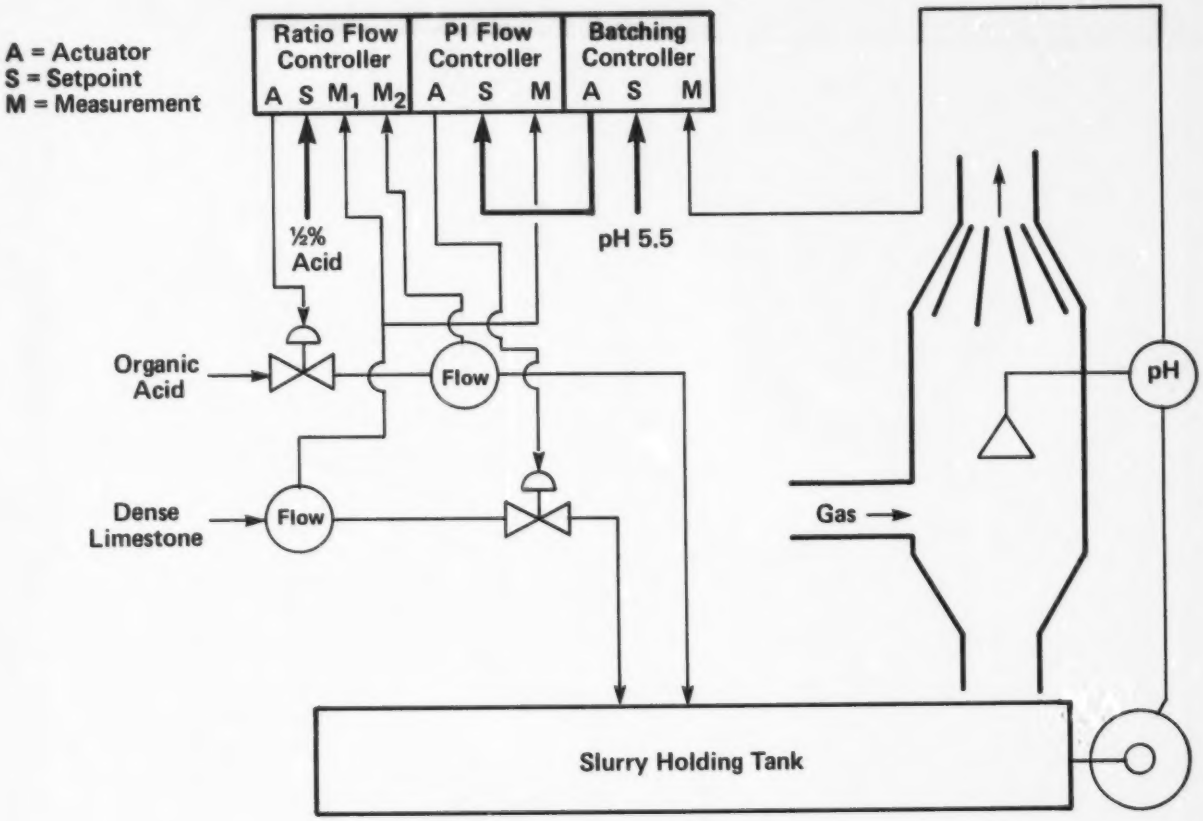

Figure 7-3. pH Feedback slurry control system with buffer (reproduced from Garrett, 1981).

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especially recommended for low slurry reaction gain operation (i.e., K_p <0.25 ΔpH/Δ gpm). The parallel combination of the two control methods is shown in Figure 7-4. In this configuration, slurry pH is the primary controlled variable. The limestone flow rate is determined by a computed Ca/SO_2 feedforward material balance. The feedforward gain (stoichiometric ratio) is adjusted by an integrating-only controller in response to differences between the set point and slurry pH values. A feedback deadband PI controller provides backup pH control in case of feedforward system malfunction or major process upsets. Finally, disturbances in the dense limestone addition may be accommodated by cascade control of the limestone flow.

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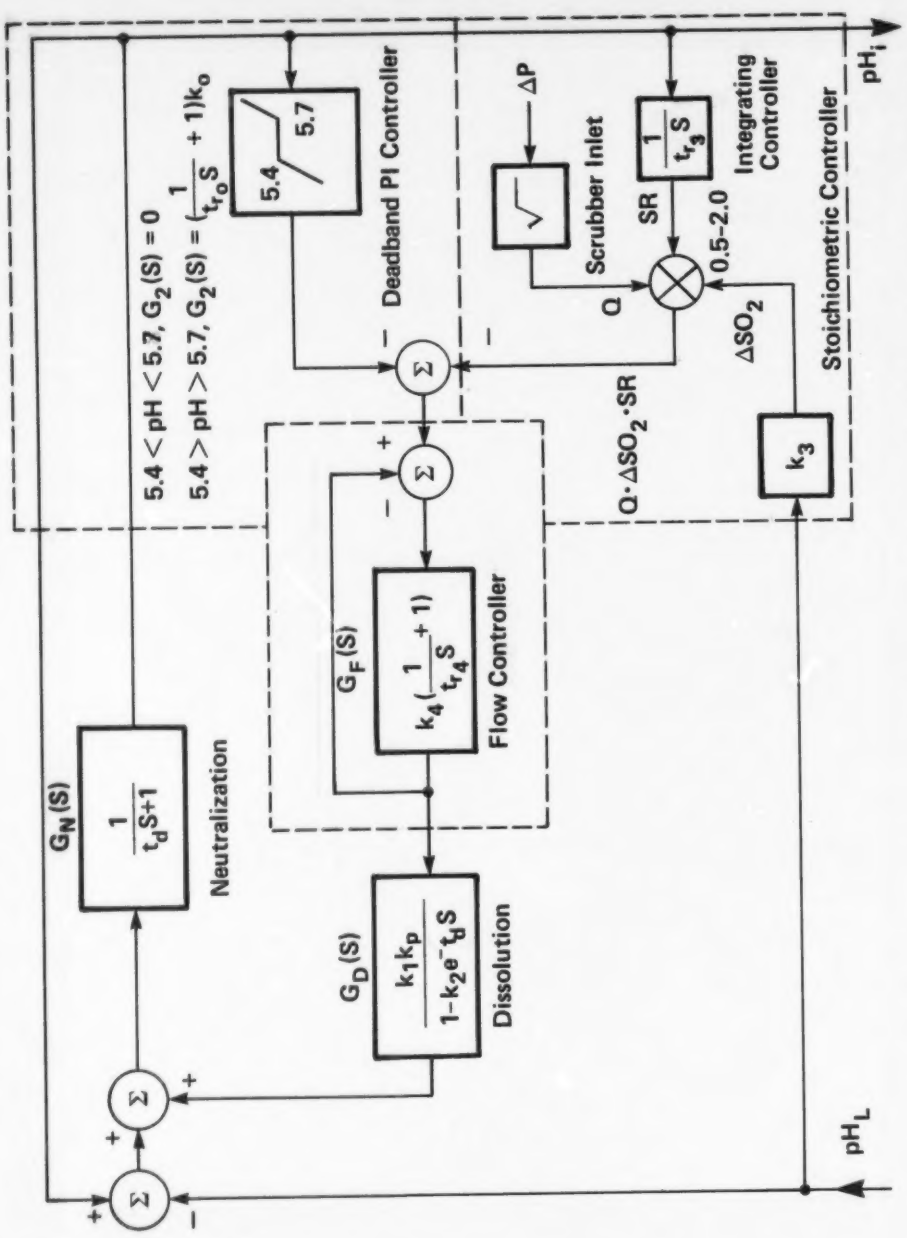

Figure 7-4. Combined stoichiometric and pH control system. (reproduced from Garrett, 1981).

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This book on classical control theory stresses both physical understanding and the utilization of simple and powerful tools for solving a wide variety of problems.

Instrument Society of America (ISA). ISA Directory of Instrumentation™, Volume 4, 1982-83.

This directory is a comprehensive catalog that supplies alphabetical listings of generic products and manufacturers.

Instrument Society of America (ISA). INTECH. (Instrumentation Technology). (Periodical).

This monthly journal of the Instrument Society of America covers new technical developments and applications in instrumentation and process control.

Luyben, W. L. Process Modeling, Simulation, and Control for Chemical Engineers, McGraw-Hill Book Company, 1973.

This text is a very readable summary of control system theory as applied to chemical engineering operations with special attention being given to mathematical modeling and computer simulation of chemical process operations.

Murrill, Paul W. Fundamentals of Process Control Theory. Instrument Society of America (ISA), 1981.

This first book in ISA's "Independent Learning Modules" series is aimed at teaching basic principles of process automation. The ILM series is designed for independent self-paced, self-study.

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APPENDIX A. ANALYSIS OF SYSTEMS DYNAMICS AND CONTROL SYSTEMS

A-1. TIME-DOMAIN ANALYSIS

The time-domain representation of control systems is the basis for modern control theory and system optimization. Analysis of systems dynamics in the time domain always involves solution of differential equations, either analytically or numerically. If the differential equations are linear (i.e., the coefficients of the time-dependent variables are constant) or can be linearized, the Laplace transform or the frequency response-Fourier transform may be applied, greatly simplifying solutions of the equations. The solution of a time-domain formulation of a control system problem is greatly facilitated by the use of computers. The availability and speed of digital computers in recent years has influenced the extent of use of time-domain analysis. The time-domain techniques may be utilized for nonlinear, time-varying and multivariable systems. A time-varying control system is a system for which one or more of the parameters of the system may vary as a function of time--for example, the number of available, active sites in a catalyst bed.

When relating results of a control system analysis to selection and application of actual system components (such as sensors, transmitters, and controllers), the time-domain solutions must be utilized. Results from Laplace- or frequency-domain analysis must be converted to the time domain since hardware performance criteria are specified in terms of time response.

The time-domain analysis and design of control systems utilizes the concept of the state of a system. The state of a system is a set of numbers such that the knowledge of these numbers and the input functions will, with the equations describing the dynamics, provide the future state and output of the system (Dorf, 1967).

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The time response of time nonlinear systems may be addressed in the time domain utilizing discrete-time approximation methods which are readily applicable to digital computer programming. Discrete time approximation is based on the division of the time axis into sufficiently small time increments, with the values of the state variables then evaluated at each successive time interval.

In some instances, useful information can be obtained about the dynamics of a system by reducing it to a steady-state equation and solving. For example, in nonlinear systems the possibility of multiple steady-states arises, one or more of which may be unstable. Evaluations by this approach are common, for example, in examining chemical reaction processes.

Reference

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Dorf, R. C. Modern Control Systems. Addison-Wesley Publishing Company, Reading, MA, 1967.

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A-2. LAPLACE DOMAIN ANALYSIS

Laplace domain techniques are used to obtain transfer functions for processes and also provide the basis for analysis of stability, an important aspect of any system. The Laplace transform method substitutes relatively easily solved algebraic equations for more difficult differential equations. The time response solution may be obtained by (1) formulating the linear differential equation, (2) deriving the Laplace transform of the differential equation, and (3) solving the resulting algebraic transform for the variable of interest. The basic definition of the Laplace transform is recalled as

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$$L[f(t)] = \int_0^\infty (f(t)e^{-st}\, dt = f(s)\,, \qquad \text{(A2-1)}$$

where s is the complex number which is the primary Laplace transform variable. The values of s lie in an s-plane, in which the two axes are real and imaginary, respectively.

Some significant properties of the Laplace transform are given below:

- The Laplace transform f(s) contains no information about the behavior of f(t) for t<0.
- The Laplace transform is linear; i.e.,

$$L\left[af_1(t) + bf_2(t)\right] = a\,L\left[f_1(t)\right] + b\,L\left[f_2(t)\right] \qquad \text{(A2-2)}$$

- A shift from s to (s-a) in the s-plane is equivalent to multiplying f(t) by e^{at}.
- For a function f(t-a), which has the value zero for $t \leq a$,

$$L[f(t-a)] = e^{-as}f(s)\,, \qquad \text{(A2-3)}$$

where f(s) is the Laplace transform of f(t).

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The significance of this theorem in control system analysis is that e^{-as} is the "delay operator" or "dead-time operator" where a is the magnitude of the delay or dead-time.

. The Laplace transform of the time derivative of a function is an algebraic expression:

$$L\frac{d\,f(t)}{dt} = s\,f(s) - [f(t)]_{t=0} \qquad \text{(A2-4)}$$

when $[f(t)]_{t=0} = 0$, and $L\frac{d\,f(t)}{dt} = s(fs)$.

. Division of f(s) by s is equivalent to integration in the time domain. That is,

$$L\int_0^t f(t)dt = \frac{1}{s}\,f(s). \qquad \text{(A2-5)}$$

In the Laplace transform method of solution, the terms of a differential equation are converted to their transforms, and the resulting algebraic expression is solved for the unknown functions. Finally, the function of t, which has the subject Laplace transform, must be found by inversion of the transform. This function will satisfy the differential equation and the initial conditions. The inversion step is frequently the most involved part of the solution procedure. It should be emphasized that only linear differential equations can be treated in the above manner. Linearization of nonlinear functions (e.g., functions involving squares, exponentials, or products of variables) can be accomplished by Taylor series expansions around the steady-state operating variable with smaller terms being neglected (Luyben, 1973).

These steps in Laplace domain analysis are illustrated in the example that follows.

Example

The behavior of a constant holdup, isothermal, constant throughput stirred-tank reactor with a first-order irreversible reaction is described by a linear first-order differential equation:

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$$\frac{dCa}{dt} + \left(\frac{1}{\tau} + k\right) C_A = \frac{1}{\tau} C_{AO} \qquad \text{(A2-6)}$$

where C_{AO} is the concentration of A in the entering stream, C_A is the concentration in the exit stream, and τ and k are constants.

Initially, the concentration in the tank, $C_A(0) = 0$. At time zero, a step change in feed concentration is made so that C_{AO} at t=0 is $\overline{C_{AO}}$.

Laplace transforming Equation (A2-6) gives:

$$sC_A(s) - C_A(0) + \left(\frac{1}{\tau} + K\right) C_A(s) = \frac{1}{\tau} C_{AO}(s) \; . \qquad \text{(A2-7)}$$

Accounting for the initial condition, $C_A(0) = 0$, grouping terms, and rearranging then yields the system transfer function, G(s).

$$G(s) = \frac{C_A(s)}{C_{AO}(s)} = \frac{1/\tau}{s + k + 1/\tau} \; . \qquad \text{(A2-8)}$$

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The denominator polynomial of the transfer function is called the characteristic equation. For physical systems, the polynomial of the numerator will be of lower order in s than will be the denominator. The roots of the denominator of the transfer function are called the poles of the transfer function. These are the values of s at which G(s) goes to infinity. The poles of the transfer function are equal to the roots of the characteristic equation. The characteristic equation of a system is its most dynamic feature. The roots of the characteristic equation will dictate if the system is fast or slow, stable or unstable, overdamped or underdamped.

Equation (A2-8) may be rearranged into a standard form as follows:

$$G(s) = \frac{\frac{1}{1 + k\tau}}{\frac{1}{k + \frac{1}{\tau}} s + 1} = \frac{K_p}{\tau_p s + 1} \qquad \text{(A2-9)}$$

where K_p is the process steady-state gain (i.e., ratio of steady-state value of output over input), and τ_p is the process time constant. The

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pole of the transfer function is the reciprocal of the process time constant.

The transfer function is a property of the system only and is applicable for any input. The form of the transfer function (in this case a first-order lag, i.e., one pole) and the numerical values of the steady-state gain and process time constant characterize the system in a very compact and usable form.

Now consider the dynamics when at time t, a step change in the magnitude of C_{A0} is imposed, i.e.,

$$C_{A0}(t) = \overline{C_{A0}}\ u(t) . \qquad \text{(A2-10)}$$

Laplace transforming Equation (A2-10) gives:

$$C_{A0}(s) = \overline{C_{A0}}\ \frac{1}{s} . \qquad \text{(A2-11)}$$

Substituting in the transfer function expression,

$$C_A(s) = \frac{k_p}{\tau_p s + 1}\ \frac{\overline{C_{A0}}}{s} . \qquad \text{(A2-12)}$$

Inversion of the transform using partial fractions yields the value of $C_A(t)$:

$$C_A(t) = K_p\ \overline{C_{A0}}\ (1 - e^{-t/\tau_p}). \qquad \text{(A2-13)}$$

Thus, the transient and steady-state performance indices may all be specified in terms of the Laplace transform functions for linear systems.

Stability

Analysis in the s-plane is particularly advantageous in evaluating the relative stability and transient response performance of closed-loop control systems. These are directly related to the location in the s-plane of the closed-loop roots of the characteristic equations. Two methods of particular importance for these evaluations are (1) the Routh-Hurwitz stability method, and (2) the root-locus method.

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The Routh-Hurwitz stability method provides a check of absolute stability by considering the characteristic equation of the system,

$$a_n s^n + a_{n-1} s^{n-1} + \ldots + a_1 s + a_o = 0 .$$

The roots of the characteristic equation must lie in the left-hand portion of the s-plane (i.e., the coefficients of the polynomial must have the same sign, and the coefficients are all nonzero); otherwise, the system is unstable. These conditions, however, do not assure that the system is stable. The Routh-Hurwitz criterion is a necessary and sufficient criterion for the stability of linear systems. If there are no roots with positive real parts, the system is stable. The Routh-Hurwitz criterion can be applied using determinants or arrays to determine if there are any roots of a characteristic equation in the right half of the s-plane.

The criterion can be applied to either open- or closed-loop systems by merely using the appropriate characteristic equation. While the criterion is quite useful, it does have definite limitations. It cannot handle systems with dead time. It tells if the system is stable or unstable, but it gives no information about how unstable or stable the system is.

To determine "relative stability," it is necessary to investigate the relative damping of each root of the characteristic equation. This relative stability may be defined as the property that is measured by the relative settling times of each root or pair of roots. Therefore, relative stability is represented by the real part of each root.

Relative stability as determined using the Nyquist criterion is discussed further in connection with frequency domain analysis.

Root Locus

The root locus technique, developed by Evans in 1948, is a graphical method for drawing the locus of roots in the s-plane as a parameter (gain, typically) is varied. The actual values of the roots of the characteristic equation are determined from the root locus, and the transient response of a system to an arbitrary forcing function can be obtained. An approximate sketch is obtained in order to analyze the initial design of a system and

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to determine suitable alterations of the system structure and the parameter values. A root locus curve is a plot of the roots of the closed-loop characteristic equation of a system as a function of the feedback controller gain.

The root-locus technique may be used to great advantage in conjunction with the Routh-Hurwitz criterion and the Newton-Raphson iteration approach for rapid polynomial roots solution.

Reference

Luyben, W. L. Process Modeling, Simulation, and Control for Chemical Engineers. McGraw-Hill Book Company, 1973.

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A-3. FREQUENCY DOMAIN ANALYSIS

Frequency response methods are a very practical and important alternative to s-plane analysis. The frequency response of a system is the steady-state response of the system to a sinusoidal input signal. The sinusoid is a unique input signal, and the resulting output signal for a linear system--as well as signals throughout the system--is sinusoidal in the steady-state; it varies from the input waveform only in amplitude and phase angle.

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One advantage of the frequency response method is that, where sinusoidal test signals are available, frequency response is easily determined and is the most reliable and uncomplicated method for the experimental analysis of a system. The second advantage is that the design of a system in the frequency domain provides the designer with control of the bandwidth of the system, and some measure of the response of the system to undesired noise and disturbances.

A third advantage of the frequency response method is that high-order, complex systems are often very easily handled because of elimination of a need to manipulate transfer functions by combining complex numbers numerically. Processes with dead time are handled easily by this method in contrast to Laplace domain analysis.

The basic disadvantage of the frequency response method for analysis and design is the indirect link between the frequency and time domains. Direct correlations are somewhat tenuous, and in practice the frequency response characteristic is usually adjusted by including various design criteria that result in a satisfactory transient response.

Frequency response methods basically apply the Fourier, rather than the Laplace, transform. The Fourier-transform pair is written:

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$$F(j\omega) = \int f(t)e^{-j\omega t}\,dt \ , \qquad \text{(A3-1)}$$

$$f(t) = \frac{1}{2\pi}\int F(j\omega)e^{j\omega t}\,d\omega \ . \qquad \text{(A3-2)}$$

It can be easily shown that the Laplace and Fourier transforms differ only in the complex variables. Thus, if the Laplace transform of a function, $F_1(t)$, is known to be $F_1(s)$, one can obtain the Fourier transform of this same function, $F_1(j\omega)$, by setting $s = j\omega$ in $F_1(s)$. The frequency response method allows one to consider the transfer function, $G(j\omega)$, and examine the amplitude and phase characteristics of the system.

The primary tools of frequency response analysis are graphical outputs, displaying: (1) the real and imaginary parts of the transfer function,

$$G(j\omega) = G(s)\big|_{s = j\omega} = R(j\omega) + jX(j\omega), \qquad \text{(A3-3)}$$

as polar plots (sometimes called Nyquist plots); and (2) Bode diagrams (logarithmic plots) of $G(j\omega)$, which display the magnitude and phase angle versus frequency. These plots for several commonly encountered transfer functions are illustrated in Figure A-1.

Bode plots are used so much because they make it easy to combine transfer functions. The complex transfer function can be broken down to its simple elements--leads, lags, and dead time. Each of these is plotted in Bode diagrams and, finally, the total complex transfer function is obtained by adding the log modulus and phase angle curves. All systems of equations can be solved very easily in the frequency domain. The procedure can be summarized as follows:

1. The ordinary differential equations are Laplace-transformed.
2. The $j\omega$ is immediately substituted for s (or the ODE's are simply Fourier-transformed).
3. A specific numerical value of frequency, ω, is chosen.
4. The resulting algebraic equations, in terms of complex variables, are solved numerically to obtain the desired transfer function relationship. These $G(j\omega)$'s will be complex numbers that are

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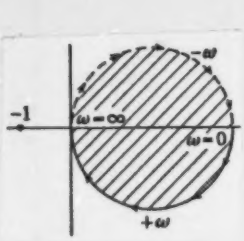
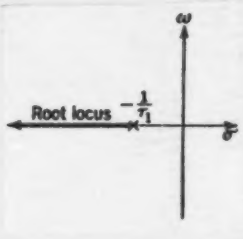
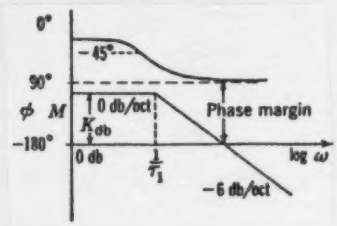
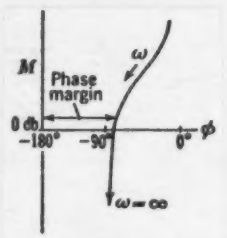
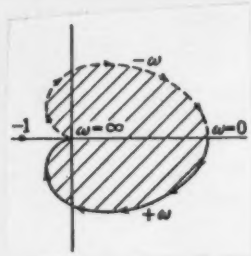
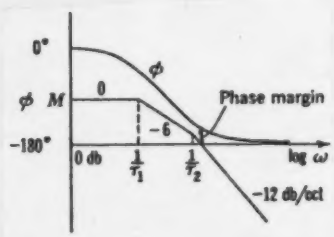
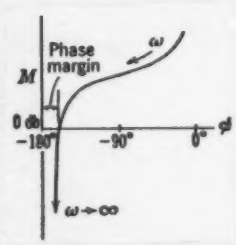
Typical transfer function; comments	Transfer function plots
$G(s) = \frac{K}{s\tau_1 + 1}$ DC amplifier; gain margin = ∞; stable.	 Polar plot  Root locus diagram  Bode diagram  Nichols diagram
$G(s) = \frac{K}{(s\tau_1 + 1)(s\tau_2 + 1)}$ Elementary, regulator; gain margin = ∞; stable.	 Polar plot Root locus diagram  Bode diagram  Nichols diagram

Figure A-1. Transfer-function plots for typical transfer function

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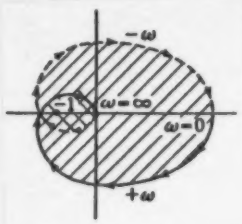
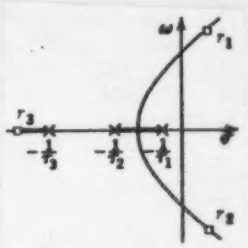
Typical transfer function; comments	Transfer function plots

$$G(s) = \frac{K}{(s\tau_1 + 1)(s\tau_2 + 1)(s\tau_3 + 1)}$$

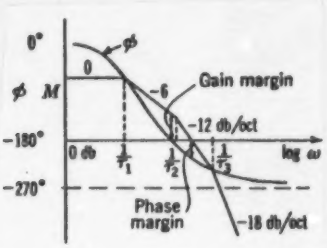
Regulator with additional energy storage component; unstable, but can be made stable by reducing gain.



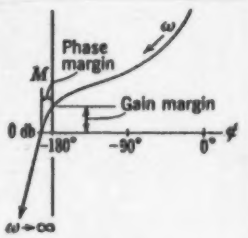
Polar plot



Root locus diagram



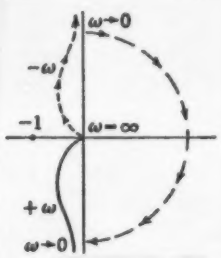
Bode diagram



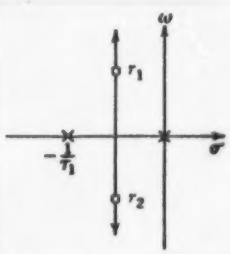
Nichols diagram

$$G(s) = \frac{K}{s(s\tau_1 + 1)}$$

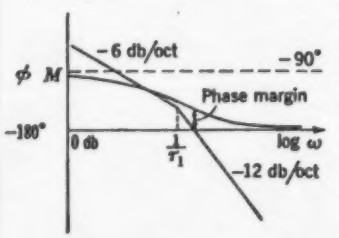
Elementary instrument servo (e.g., ac-motor, two-phase field control); gain margin = ∞; inherently stable.



Polar plot



Root locus diagram



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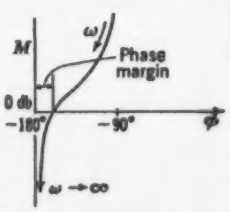

Nichols diagram

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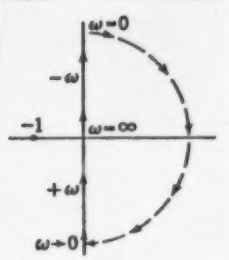
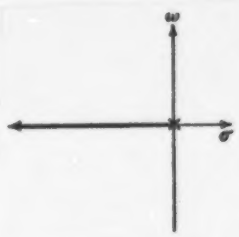
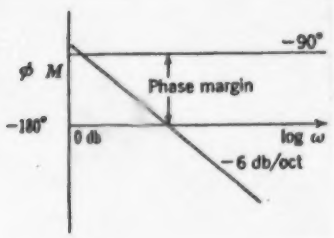
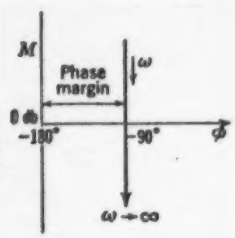
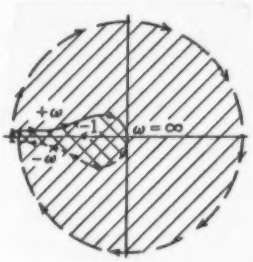
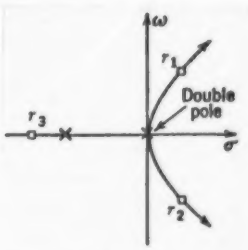
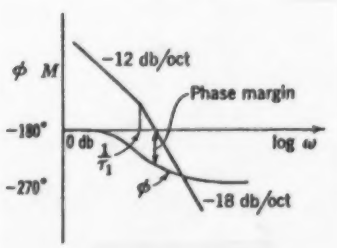
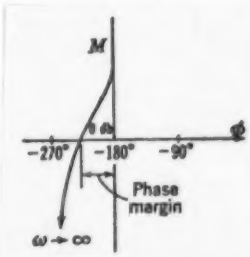
Typical transfer function; comments	Transfer function plots
$G(s) = \frac{K}{s}$ Ideal integrator; stable.	 Polar plot  Root locus diagram  Bode diagram  Nichols diagram
$G(s) = \frac{K}{s^2(s\tau_1 + 1)}$ Inherently unstable; must be compensated.	 Polar plot  Root locus diagram  Bode diagram  Nichols diagram

Figure A-1 (continued)

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points on the polar or Bode plots corresponding to the specific frequency chosen.

5. Another numerical value of ω is specified, and step 4 is repeated.

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Selecting a number of frequencies over the range of interest for the process gives the complete frequency response curves. Not only will frequency response be obtained, but also several performance specifications in the frequency domain may be calculated, including the maximum magnitude, M_p, and the resonant frequency, ω_r. The Bode diagrams may be related to system error constants.

Finally, frequency domain analysis may be applied to examine the stability of a system utilizing the "Nyquist stability criterion." Application of this criterion involves mapping of contours in the complex plane and is an important application of the theory of "conformal mapping." Use of the Nyquist stability criterion allows the closed loop stability or instability to be deduced from the frequency response plots of the open-loop transfer function of the system. Stable and unstable systems are illustrated in the Nyquist plot shown in Figure A-2 (see Luyben, 1973).

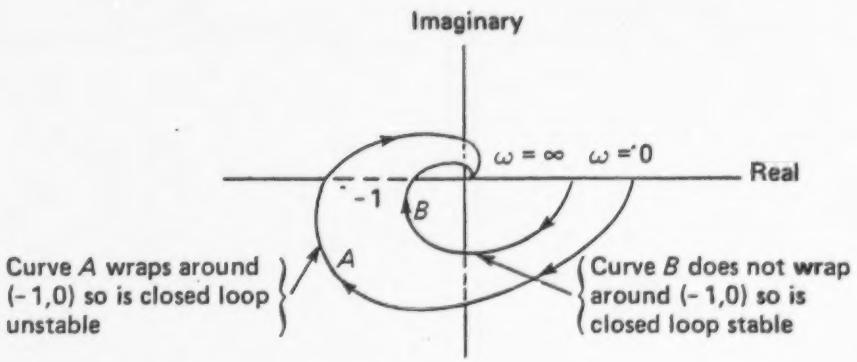

Figure A-2. Plots showing closed-loop stability or instability based on Nyquist criterion.

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The Nyquist criterion also provides the relative stability of a system as measured by the "gain margin" and the "phase margin." Gain margin is the factor by which the system gain would have to be increased to become unstable. Phase margin measures the additional phase lag that is required for the system to become unstable.

The magnitude and phase of a closed-loop system can be determined from the frequency response of the open-loop transfer function by utilizing constant-magnitude-and-phase circles on the polar plot. Alternatively, one may utilize a log magnitude-phase diagram with closed-loop magnitude and phase superimposed (called the Nichols chart) to obtain the closed-loop frequency response. In the Nichols plot, the log modulus is plotted versus the phase angle with frequency as parameter along the curve. Another measure of relative stability, the maximum magnitude of the closed-loop frequency response, M_p, is available from the Nichols chart. The Nichols diagrams are also shown in Figure A-1.

References

Luyben, W. L. Process Modeling, Simulation, and Control for Chemical Engineers. McGraw-Hill Book Company, 1973.

Dorf, R. C. Modern Control Systems. Addison-Wesley Publishing Company, Reading, MA, 1967.

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A-4. OTHER SOLUTION APPROACHES

Direct or numerical solution of differential equations (often through digital simulation) is an increasingly important technique in analyzing control systems. The approach of experimentation is also recognized--especially the advantages of sinusoidal input for solution in the frequency response domain. Dynamic behavior can be found from experimental tests, which are sometimes used when the process is thought to be too complex to model from first principles. Experimentation may also be used to find the values of some parameters that are unknown. Another important use of dynamic experiments is to confirm the prediction of a theoretical mathematical model. Besides sine wave testing, the techniques of impulse- and step-input testing are common. The transfer functions and other system response information may be directly obtained from the data gained from such testing.

Finally, the utilization of a computer for analog simulation makes use of the fact that the equations describing many electrical devices are completely analogous to the equations describing other systems--mechanical, thermal, chemical, hydraulic, etc. Basically, the analog computer simulation is accomplished by wiring or patching together electronic boxes that perform the various mathematical operations--such as addition, multiplication, integration, and differentiation--approximating the differential equation describing the the physical system. The complexity of nonlinearity of the describing equations is no obstacle for analog computer simulation. With proper gains, the analog computer can actually simulate the dynamic response of the system in real-time.

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A-5. RESPONSE TO FORCING FUNCTIONS

Figure A-3 shows the response of various systems to common forcing functions. Real systems can be approximated by the pure systems (e.g., integrator, first-order lag, cascaded first-order lags, etc.) or their combinations (Buckley, 1964). Response in the time domain $\theta_o(t)$, and in the Laplace domain, $\theta_o(s)$, are indicated in the figure. Initial conditions are assumed to be zero.

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Reference

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Buckley, P.S. Techniques of Process Control. John Wiley and Sons, 1964.

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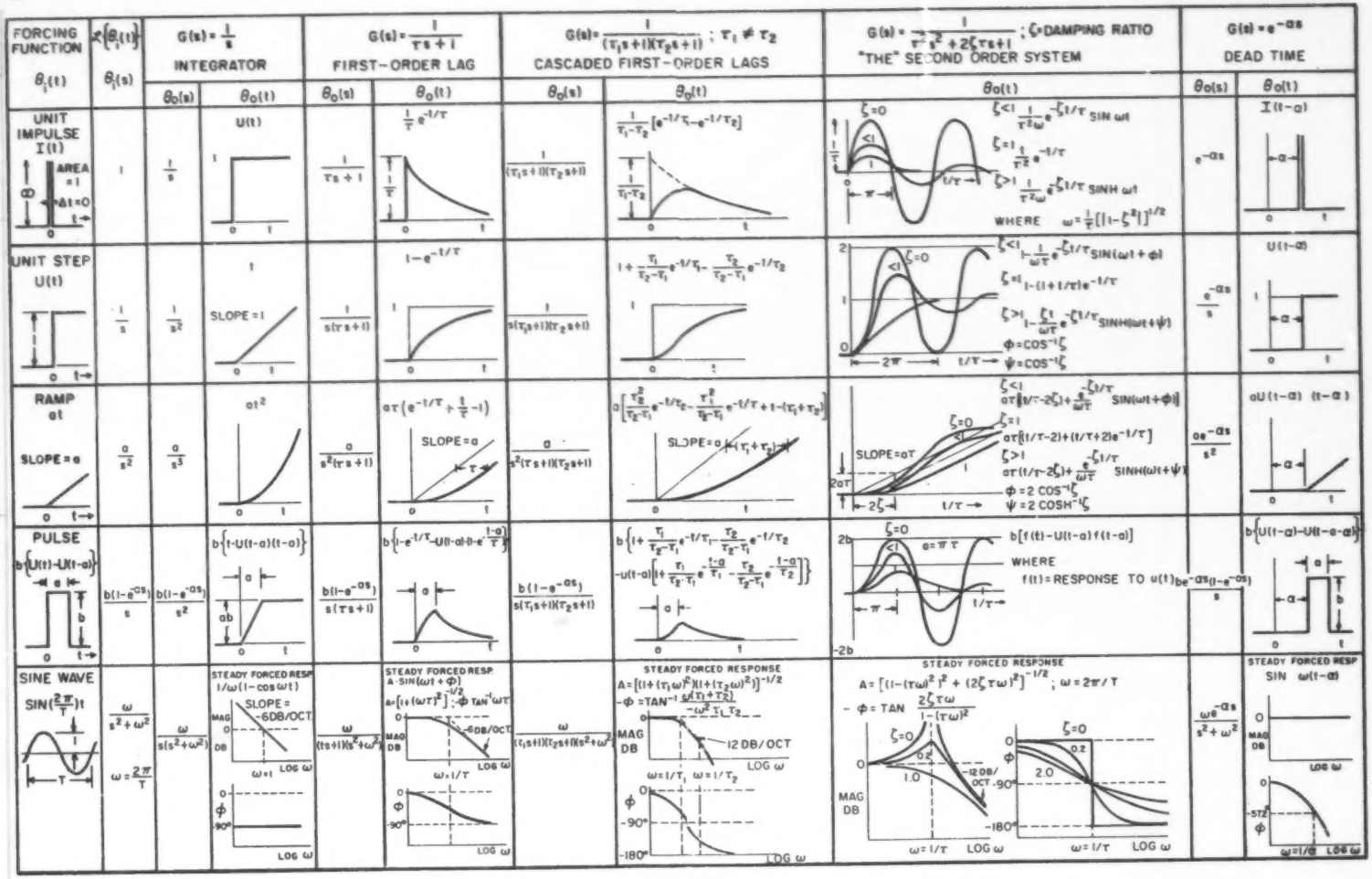

Figure A-3. Response of typical systems to common forcing functions. (reproduced from Buckley, 1964, with permission from the publisher, John Wiley and Sons, Inc.)

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A-6. AUTOMATIC CONTROL SYSTEM DESIGN AND COMPENSATION

Consideration of dynamics should be factored into the design of a plant or an environmental control system at an early stage, preferably during pilot-plant design and operation. It is often inexpensive in the early stages of the project to design a piece of process equipment so that it is easy to control. Some general guidelines and recommendations for design suggested by Luyben (1973) are listed below:

1. Use feedforward control to compensate for large, frequent disturbances.
2. Avoid lags and dead times in feedback loops.
3. Use "averaging-level" control (proportional-only controllers) to maintain inventories in surge columns.
4. Eliminate major disturbances by using cascade control systems where possible.
5. Avoid control loop interaction where possible, or use interaction compensation.
6. Check for dynamic problems at startup, and during low throughput operations.

Shown in Table A-1 are some additional general guidelines for matching the control system to the needs of the process. The table is based on conclusions by Myron (1972).

Considerable research in recent years has been directed toward identification of performance online, e.g., evaluation of parameters by verification of a mathematical model. By continuously generating a dynamic model of the process, any change in the process with time is detected. This information is then used to retune controllers to achieve some desired performance. A number of applications of this "adaptive control" have been reported in the aerospace field. However, few industrial processes have changes that occur so rapidly that such online methods are justified.

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TABLE A-1. CONTROL SYSTEM APPLICATION CHART[a]

Type of system	Process capacity	Process deadtime	Magnitude of load change	Frequency of load change	Comments
Feedback	Small or large	Small	Small	Low	
Cascade	Large	Small	Small	Low	One load variable amenable to control
Feedforward/ feedback	Large	Large	Large	High	Excellent choice at end of processing train
Adaptive	Can be applied in conjunction with any of above--usually applied to those processes where controller tuning can offset variable gain element(s) elsewhere in the control loop.				
Inferential	Should be considered when a measurement of the controlled variable is not available.				

[a]Adapted from Myron (1972).

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A suitable control system must be stable and provide an acceptable response to input commands, have low sensitivity to system parameter changes, produce a minimum steady-state error for input command, and, finally, be able to eliminate the effect of undesirable disturbances. The designer of a control system is concerned with the arrangement or the plan of the system structure and the selection of suitable components and parameters. Frequently, conflicting sets of requirements are encountered if one desires a set of performance indices or other measures to be less than specified values. For example, if it is desirable to have a percent overshoot of less than 20 percent and a normalized peak time, $\omega_n T_p = 3.3$, a conflicting requirement on the system damping ratio, q, is obtained. If the two performance requirements cannot be relaxed, the system must be altered in some way. Often the alteration or adjustment of a control system, in order to provide a suitable performance, is called compensation.

Compensation in process instrumentation is defined as "provision of a special construction, a supplemental device, circuit, or special materials to counteract sources of error due to variations in specified operating conditions" (ISA, 1979). The compensating device, often called a compensator, may be electrical, mechanical, hydraulic, or another type of device or network. A compensator "converts a signal into a function of it which, either alone or in combination with other signals, directs the final controlling element to reduce deviations in the directly controlled variable" variable" (ISA, 1979).

The selection of the compensation scheme depends upon consideration of the specifications, the power levels of various signal modes of the system, and the networks available for use. One possible division of types of compensation considers four different approaches:

1. Cascade compensation,
2. Feedback compensation,
3. Output or load compensation,
4. Input compensation.

These are all represented in Figure A-4.

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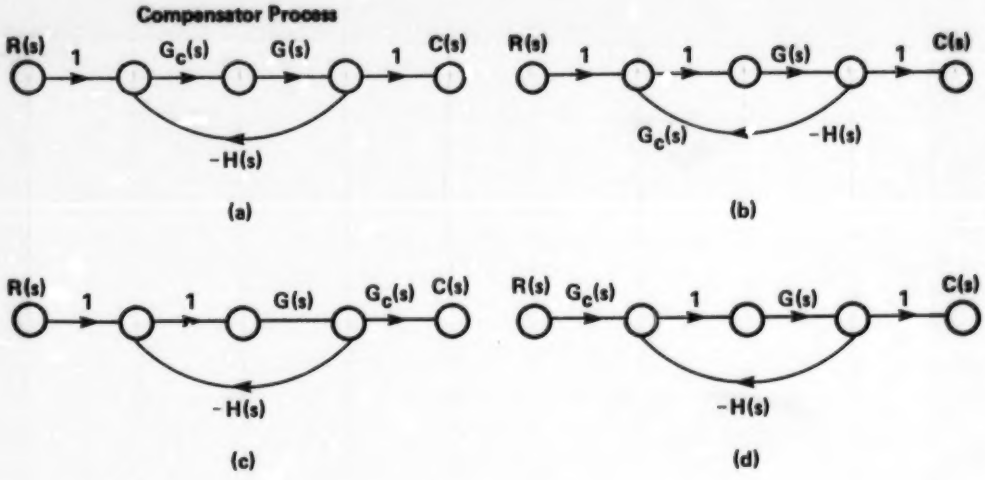

Figure A-4. Types of compensation: (a) cascade compensation; (b) feedback compensation; (c) output or load compensation; (d) input compensation.

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The transfer function of the compensator is designated as

$$G_C(s) = E_{out}(s)/E_{in}(s) \qquad \text{(A6-1)}$$

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where E represents the actuating signal.

Cascade compensation networks, either phase-lead or phase-lag, alter the shape of the root-locus, or frequency response of a system. Compensation networks are also often distinguished by their phase angle, θ, (for G_C). The phase angle is the angular difference between the input and output. If the output lags the input, θ is negative; if the output leads the input, θ is positive. For a first-order compensator, with a phase-lead network, $\theta = +90°$; and for first-order phase-lag, $\theta = -90°$.

Compensation may be studied by using a phase-lead, s-plane network on the Bode diagram, and the root-locus, s-plane diagram successively. A phase-lead compensator increases the phase margin of the system and thus provides additional stability. Phase-lead compensators should be designed on a Bode diagram when the design specifications include an error constant, and on the s-plane when design specifications, for example, include the settling time and overshoot for a step-input. When large error constants are specified for feedback systems, it is usually easier to compensate the system by using phase-lag (integration) networks. Phase-lead compensation increases the system bandwidth while phase-lag compensation decreases it. This may be an important factor when noise is present at the input or is generated within the system. Diagrams applicable to these types of analyses are shown in Figure A-1.

References

Instrument Society of America (ISA). Standard Process Instrumentation Terminology. Approved by American National Standards Institute, December 28, 1979. ANSI/ISA-S51.1, 1979.

Luyben, W. L. Process Modeling, Simulation, and Control for Chemical Engineers. McGraw-Hill Book Company, 1973.

Myron, T. J., Jr. Guidelines for Control System Design. Instrumentation Technology, January 1972.

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